Pulverized-Fuel Combustion in Trouble

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Pulverized-fuel (PF) firing is the combustion technique used in all of our great power-generating stations based upon coal. Until recently, an aura of "inevitability" surrounded this technique and tended to protect it from competitive ideas. Now, almost overnight, a sharp increase in concern over environmental quality has placed the future of PF firing in doubt. A number of groups, some represented at this Symposium, are exploring alternative procedures which, first, promise to reduce the cost of coal-fired power, and, second, offer opportunities for reduced cost of control of both ash and sulfur oxides emissions.

At least two concrete commercial developments are in the offing which should go far to dispel PF combustion's aura of inevitability:

- the installation of an Ignifluid boiler in the anthracite district of Northeastern Pennsylvania;
- Lurgi's installation of a combined gas- and steam-turbine power unit incorporating pressure gasification of coal.

Experimental and design studies now in progress also point toward new paths for coal development:

- work on the fluidized-bed boiler;
- interest shown by firms catering to the power industry in studies of combined-cycle arrangements generating power from gas produced from coal;
- continued interest in possibilities for use of coal in fuel cells and magnetohydrodynamic (MHD) devices;
- work directed toward development of a "Coalplex" yielding pipeline gas or liquid fuel or chemicals and low-sulfur coke for power use.

Viewed altogether, these commercial and experimental activities lead to the inescapable impression that a revolution in coal-power practice may be at hand.

The National Air Pollution Control Administration (NAPCA) has recognized the opportunity to steer this revolution into paths leading to better ways to control ash and sulfur emissions.

NAPCA has engaged Westinghouse Electric Co. to direct the development of a non-polluting fluidized-bed boiler, and United Aircraft to study schemes for generating power from clean gas made from coal.

A technique as mature as PF firing is hard to displace, and its advocates can be expected to work hard to keep it viable. It may be useful to review briefly the problems which they will face in an historical context.

The concerns of students of combustion may be listed roughly in the order in which they have arisen:

- to burn coal with an acceptably small loss of carbon to smoke and ashes;
- to provide clean combustion gases suitable for heating materials liable to be spoiled by ashes;
- to provide combustion gases for discharge to a stack which were sufficiently free of grit as not to constitute a neighborhood nuisance;
- to burn coal at the large throughputs needed to generate electricity after about 1925;
- to provide stack gases to meet increasingly higher standards for content of fly ash;
- to provide stack gases low in sulfur oxides.

Until about 1895, all devices for burning solid fuel handled the fuel in a bed at rest. In some, the bed gravitated downward in a shaft. In others, the bed rested or moved horizontally on a grate. Steam power engineers developed ingenious devices, the first patented by Watt himself in 1785, to feed coal continuously to a bed on a grate and to discharge ashes. Grit emissions from some of the grate devices were small, although their designers were at first more concerned with limiting losses of carbon.

The advantages of dealing with the coal in several steps were appreciated early. From 1735 onward, ironmakers in England used coke from beehive coke ovens. After 1800 an industry arose to supply illuminating gas, marketing coke as a byproduct. In 1836 gas producers were introduced to derive from coke a dust-free fuel gas suitable for burning where cleanliness was desired.

Often, a major incentive to technical change has been growth of demand for a commodity, making obsolete a technique whose scale-up to large size is difficult or uncertain. By the 1890's, cement manufacturers felt need for equipment of larger capacity than the shaft kilns used hitherto. An attempt to operate a rotary kiln for cement-making with producer gas was a failure. A kiln was operated satisfactorily with petroleum, but this fuel was then too expensive. The experience suggested that a suitable flame might be sustained by injecting pulverized coal into a rotary kiln via an air blast from a nozzle. Shortly, the cement industry developed techniques for pulverizing coal and burning the coal powder. Edison participated in this work, attesting its importance to late 19th-century technology.

By about 1915, steam power engineers realized that electricity demands would soon require steam flows larger than could be conveniently provided by a grate-firing technique. They felt an acute need for a new combustion procedure easier to scale upward in size than the existing grate-combustion devices. The experience of the cement industry was at hand: coal pulverizers, coal-conveying systems, and PF firing nozzles were available on the market. Engineers found it relatively inexpensive to undertake experiments on PF firing for raising steam. The work led to the Lakeside Station in Milwaukee. After the commissioning of two 20,000-Kw turbines in this station in 1922, PF firing soon became the choice for nearly all new power-station construction.

Engineers of the day regarded the PF boiler to be an advance from standpoint of dust emission. Herington (1) wrote in 1920:

> "It is quite true that perhaps 60 per cent of the ash goes up through the stack. This ash is of such light flocculent nature that it is dissipated over a wide area before precipitation occurs and no trouble can be expected from this source, although the amount of tonnage put out through the stack per day seems great. This is proved by the 'Lopulco' installation [at Oneida Street Plant of Milwaukee Electric Railway & Light Co.] where, after a period of two years' operation, although the plant is located in the heart of the business district of Milwaukee, no complaint has been heard from this source and no evidence of any ash or dust can be found on the roofs of any of the buildings in the vicinity. It is quite possible that this dust is of such fineness and such a nature that it is not precipitated until it encounters moisture."

It would appear that the engineer of 1920 was more concerned for his immediate neighbors than for a city or a region. He soon heard about it if a nearby housewife found "soot" on her wash, but voices were not yet raised concerning insults to lung tissue by fine matter. Would PF firing have seemed attractive for development if engineers had felt something like today's concern about fly ash?

A dry-bottom furnace, having steeply sloping walls, allows about 80% of the ash to leave with the gases, while the remainder drops out of the bottom in solid form. A wet-bottom furnace has a relatively flatter bottom and retains ash for a much longer time, so that about one-half leaves as molten slag. A cyclone furnace uses a coarser grind of coal and burns the coal in an intense combustion zone in which coal and gases whirl in cyclonic fashion. The effect is to separate about 70 to 90% of the coal's ash as a slag which can be tapped from the bottom. The changing attitude toward dust emissions is illustrated by the claim advanced in the 1930's, when the cyclone furnace was introduced, that it substantially solved the emission problem.

Figure 1, after Ramsdell and Soutar (2), illustrates the growth in concern over dust emissions. For more than 10 years, Consolidated Edison Co. of New York has recognized that the metropolitan settings of its stations imposes the necessity to provide equipment collecting fly ash at an efficiency greater than 99%. This necessity has led to electrostatic precipitators of great size, such as the one at the 1000-Mw unit of Con Edison's Ravenswood Station. This is shown schematically together with the boiler in Figure 2. There are two banks of precipitators, each 58 x 230 feet in plan and 75 feet in height. The enclosed volume is more than three times greater than the two combustion chambers of the Ravenswood unit, each 34 x 64 feet in plan and 138 feet in height. The Ravenswood precipitator cost $10,000,000 -- i.e., $10 per kilowatt. It has provided a collection efficiency of 99.5% in tests.

The Ravenswood precipitator operates at 700°F, while earlier precipitators in Con Edison's system generally operated at around 300°F. A reason for the higher temperature, which needs a larger precipitator to achieve comparable performance, was the introduction of coals of below 1.0% sulfur into Con Edison's system. Because ash from low-sulfur coal displays a high electrical resistivity at 300°F, a precipitator for this coal and this temperature would have to be much larger than a precipitator for a high-sulfur coal in any case, as Figure 3 shows (2). Figure 4 illustrates the rising cost of dust collection over the years, paralleling the increase in dust-collection efficiency (2).

Few existing coal-fired stations are equipped with precipitators of such high efficiency as those in Con Edison's system. In future PF stations, the power industry may find it hard to escape a cost on the order of that incurred at Ravenswood for fly ash control. A trend may be in the making, exemplified by the projected Four Corners Station in Arizona, toward scrubbing for fly ash recovery, in the hope that the costs of fly ash and sulfur oxides control may be shared.

A major drawback of PF firing for the future lies in the fact that a simple, one-step combustion places the coal's sulfur promptly into a form difficult to collect and recover. For typical coals, the combustion gases contain about 0.2 to 0.3% SO_2 by volume. The Ravenswood precipitator handles 4.3×10^6 cubic feet of gas per minute. The chemical treatment of such a vast throughput for removal of a constituent present in such small amount is almost certain to be costly.

Since the 1930's, research and development teams have worked upon many ingenious ideas for capturing SO_2 in stack gases from PF boilers. The history of many of these efforts is depressing: initial enthusiasm followed by abandonment when the economic facts became clear. At the moment, some half-dozen or so schemes are "alive", but none has passed the hurdle of commercial operation at the several-hundred-Mw scale of power generation common in the United States.

Recently, some argument, primarily semantic, has arisen concerning the "commercial availability" of systems for SO_2 control. Normal business prudence would argue against putting in a large number of several-hundred-Mw installations, simultaneously, for any of the now-available systems. An over-enthusiastic heralding of these systems could lead to pressure for such installations from environmentalists not overly concerned with either business or technological considerations. If the pressure succeeds, so much money and hope would be committed to the installations that funding for development work on more advanced schemes for sulfur oxides control would be difficult to obtain.

The history of classic disasters of engineering -- post-War Fischer-Tropsch synthesis, fluid hydroforming, nuclear-powered flight, numerous advanced-design aircraft, and more recently, Oyster Creek and high-speed rail equipment -- should teach prudence in the application of new processes on a giant scale. Many such disasters are a result of too-rapid application to meet an urgently felt need.

If trouble should develop almost simultaneously in a number of stack-gas cleaning installations, the news would reinforce the already general belief that pollutants from coal combustion are "impossible" to control, and might contribute toward another round of nuclear plant construction. The danger would be especially great if development of alternatives were not already well advanced.

Schemes to control sulfur from PF combustion have a make-shift, tacked-on aspect. The time is at hand to rethink the problem of burning coal with air pollution as an early consideration.

We have already remarked that PF combustion might not have seemed so attractive to the engineer of 1920 if he had been as much concerned with fly ash as with grit. Instead, he might well have concentrated upon ways to increase the burning capacity of his familiar grate devices.

An idea was at hand. Winkler filed his historic patent for a fluidized-bed coal gasification apparatus in 1922, and its commercial use began in 1926. It does not detract from the simple beauty of the idea to fluidize a bed of coal on a travelling grate to wonder why no one came forward with this idea before Albert Godel thought of it in the late 1940's. The "inevitability" of the PF technique was too inhibiting. Godel has stated that he himself did not at first conceive that his Ignifluid system might go into large utility boilers, and he believes he lost many years for lack of this concept.

Figures 5 and 6 give cross-sectional views through the lower portion of Godel's Ignifluid boiler (3). Godel has found that the ash of substantially all coals is self-adhering at a temperature in the vicinity of 2,000°F, no matter how much higher the ASTM ash-softening temperature may be. Coal is supplied in sizes up to 3/4 inch. As a coal particle burns, ash is released. Ash sticks to ash and not to coal, and ash agglomerates form. They sink to the grate, which carries them to the ash pit. Godel's bed operates adiabatically, except for radiation from the upper surface. The bed is rich in carbon, and combustion is incomplete within the bed. Secondary air, admitted over the bed, completes the combustion.

As a result of the high fluidizing-gas velocity (about 10 feet per second) and low air-to-fuel ratio, the coal-treating capacity of Godel's travelling grate is roughly 10 times greater than that of previous grate-combustion devices.

Recently, Babcock-Atlantique has promoted use of the Ignifluid boiler in large stations (4). A 60-Mw unit is in operation at Casablanca, and negotiations are well advanced for a 275-Mw unit to burn and remove accumulations of anthracite waste in Northeastern Pennsylvania. The waste has a high ash content, and Godel's system is uniquely capable of dealing with it.

For nearly 30 years, various groups have attempted, without much success, to burn pulverized fuel at high pressure to furnish hot gases to drive a gas turbine. The work to be reported here by BCURA and Lurgi point to paths of development whereby coal may take advantage of the substantial cost reductions which combined-cycle operation can afford.

As United Aircraft will report, the inevitable advance in gas temperatures for gas-turbine operation will bring an incentive to increase the power output from the gas turbine of a combined-cycle operation to levels of 50% and beyond (5). These developments will create an incentive to find techniques for gasifying coal in systems of high capacity and efficiency. For the American power industry, a gasifier handling the coal for 1,000-Mw in a single unit, or at most a few units, represents a reasonable target of development.

Fluidization at high velocity, perhaps with use of Lurgi's "circulating fluid bed" technique (6), comes immediately into mind.

There may be a way to combine this technique with ash agglomeration, for example, as practiced by Jéquier and collaborators at CERCHAR (7, 8).

Suppression of sulfur oxides from a two-step combustion of coal at high pressure should be far easier than from PF combustion. Sulfur would be available as H_2S, present in a far smaller volume flow of gas.

Finally, I call attention to the arrangements which have been made to bring liquefied natural gas from abroad, at prices which bring sharply into view the alternative of converting volatile matter in coal into synthetic gas. This development lends urgency to studies of schemes like the "Coalplex" depicted broadly in Figure 7. Much work sponsored in recent years by the U.S. Office of Coal Research has been directed toward development of such a Coalplex, especially work by Consolidation Coal Co. and FMC Corp.

The appearance of Coalplexes will result in availability of large supplies of low-sulfur coke, for which PF combustion is poorly suited. This fact is a powerful incentive to ready a better technique for combustion of carbon.

Figure 8 depicts broadly a logical precursor to the Coalplex of Figure 7 (9). This scheme would generate baseload power from the combustion of volatile matter, and would ship low-sulfur coke to power stations at a distance.

We see a natural evolution:

- The first Coalplex would be justified simply for its economy in dealing with sulfur.
- Later, modifications would "cream off" limited amounts of pipeline gas or liquid from volatile matter. Simplicities in the processing of volatile matter to products of higher value would result from opportunity to throw off high-level waste heat to steam for power.
- As time passed, further modifications would expand production of gas or liquid.

Ultimately, the recovery of sulfur from coal would be viewed as a mere incidental.

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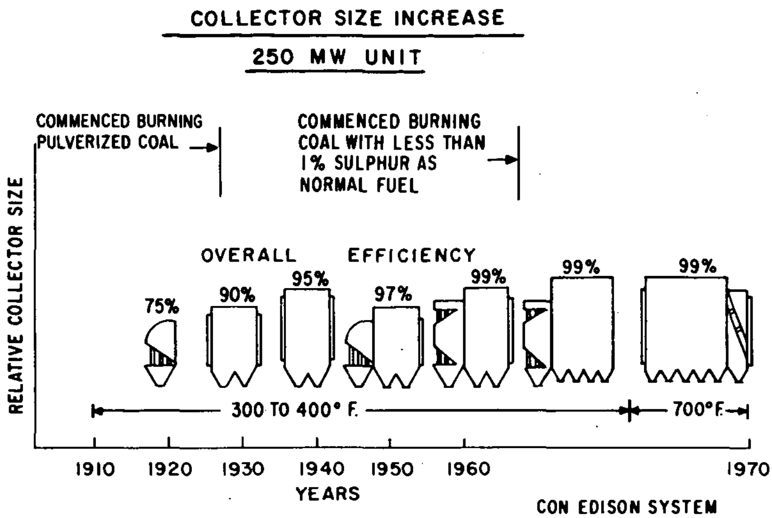

Fig. 1. Increase in size of dust collection equipment in Con Edison system, after Ramsdell and Soutar (2).

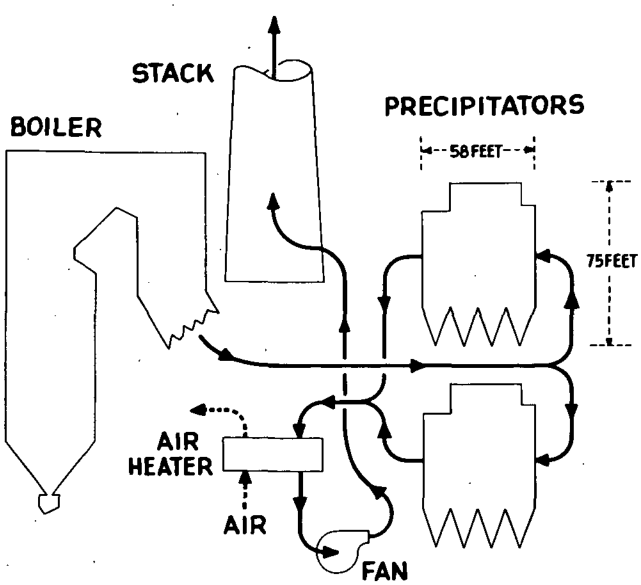

Fig. 2. Schematic view of boiler and electrostatic precipitators of 1000-Mw unit at Con Edison's Ravenswood Station.

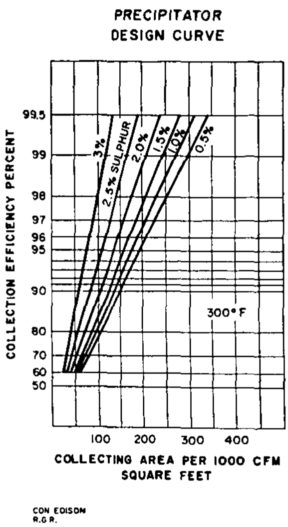

Fig. 3.

Relationship between collecting area and collection efficiency, as function of sulfur content of coal, for operation at 300°F, after Ramsdell and Soutar (2).

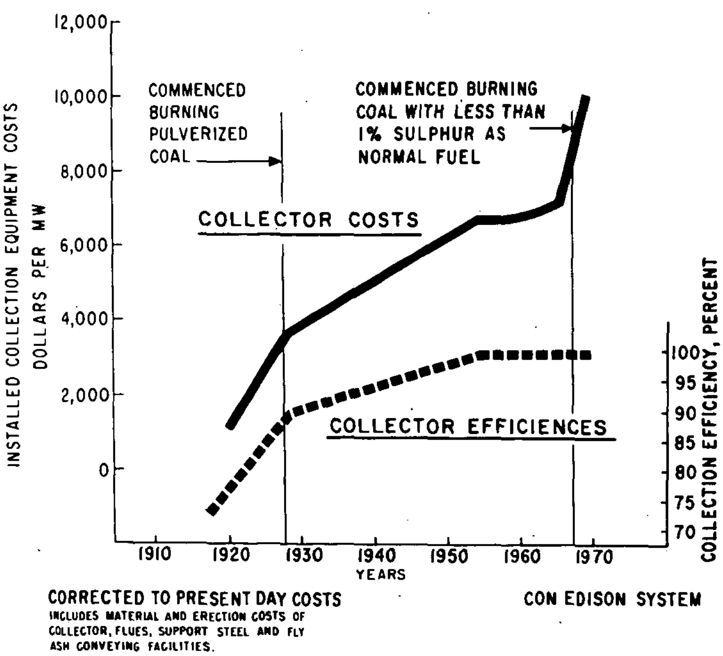

Fig. 4. Increase in cost of dust collecting equipment in Con Edison system, after Ramsdell and Soutar (2).

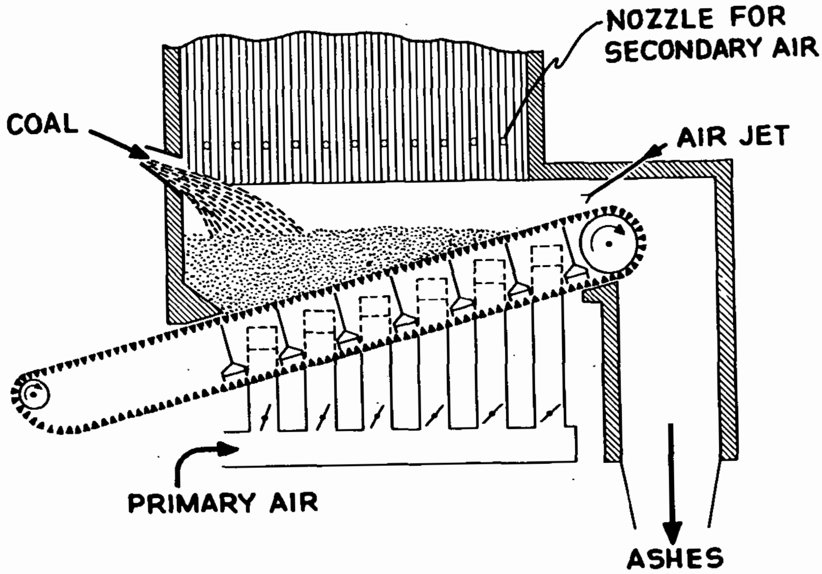

Fig. 5. Cross-sectional view of lower part of Ignifluid boiler, developed by Albert Godel and Babcock-Atlantique.

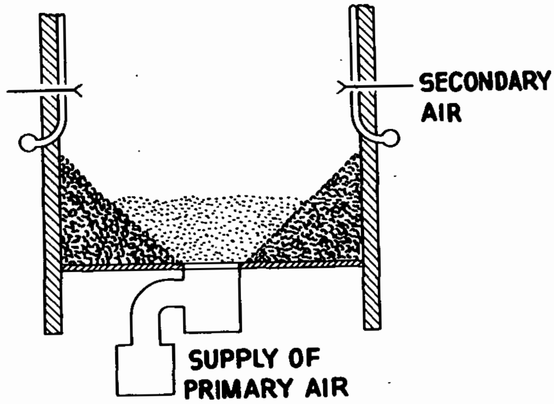

Fig. 6. Sectional view of Ignifluid boiler across the travelling grate, showing the fluidized combustion bed between two banks of static coal.

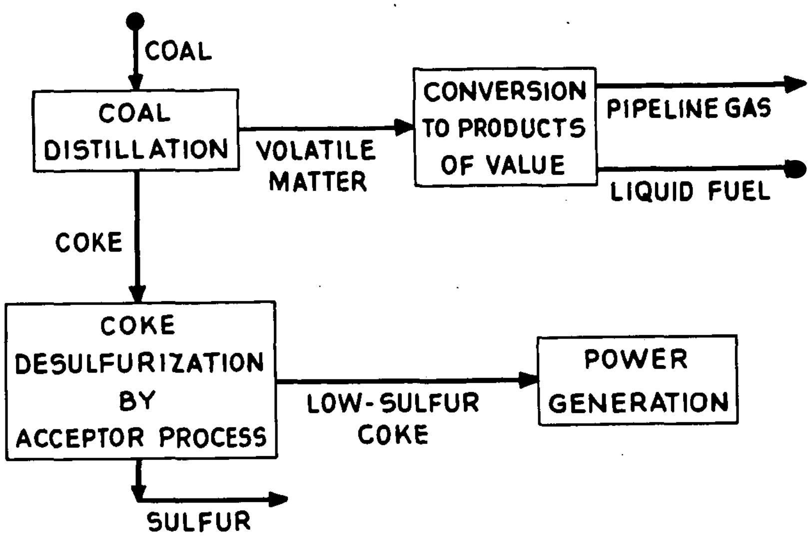

Fig. 7. A "Coalplex".

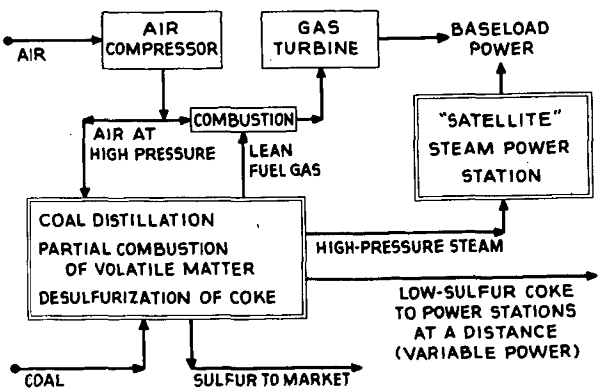

Fig. 8. Concept for a pioneering Coalplex directed toward recovery of sulfur and generation of coal power.

NEW FOSSIL-FUELED POWER PLANT PROCESS BASED ON LURGI PRESSURE GASIFICATION OF COAL

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I. NEW GAS TURBINE POWER PLANT USING LURGI PRESSURE GASIFICATION OF COAL

A new type power plant has been developed by the combination of Lurgi pressure gasification with a gas turbine process, which is capable of solving major problems in power plant technology.

The use of Lurgi pressure gasification of coal ahead of thermal power plants was proposed already many years ago, but the necessary process scheme could be realized only now after industrial gas turbines had been sufficiently developed and proved successful on a large scale.

Steinkohlen-Elektrizität AG (STEAG) undertook to work out a combined scheme of coal pressure gasification, gas turbine and steam power plant and to further improve this scheme in cooperation with their partners. The results were so promising that STEAG decided to realize this scheme and to place the order for the construction of a power plant integrated with coal pressure gasification for an output of 170 megawatts. The plant will be installed in the Kellermann power station at Lünen (West Germany) and is scheduled to go on stream by mid-1971. It shall serve chiefly for covering peak load requirements.

The present paper deals with this new process scheme.

II. REQUIREMENTS FOR COAL GASIFICATION WHEN USED IN GAS TURBINE PROCESSES

The commercial application of a gasification process in conjunction with power plants is new. The paper therefore describes first the relationship between the technology of gasification and the special features of the gas turbine process.

In the open cycle gas turbine process with internal combustion, which is used in the present case, fuel gas is burnt under pressure with a surplus of air, and the combustion gas is utilized as driving energy in a gas turbine. The fuel gas is generated by the gasification of coal. This means that the combustion reaction underlying all thermal power plants, namely

$$C + O_2 = CO_2 + 173,000 \text{ BTU/lb. mol}$$

is split up as follows:

a) Gasification of solid fuel, for instance, according to the simplest gasification reaction

$$C + \frac{1}{2} O_2 = CO + 51,000 \text{ BTU/lb. mol}$$

b) Combustion of the product gas in the combustion chamber of the gas turbine under pressure

$$CO + \frac{1}{2} O_2 = CO_2 + 122,000 \text{ BTU/lb. mol}$$

The total of these reactions is again the combustion formula. As the gas must be supplied under pressure, the first requirement demanded from the gasification process is:

1st Condition: Gasification under pressure.

In the gas turbine process, the same as in any other thermal power process, the difference between inlet temperature and outlet temperature of the gas during its depressurization is the measure for the heat rate. The higher the inlet temperature of the gas the higher the process efficiency. The maximum inlet temperature is limited by the service life of the blade material. It is therefore a requirement that the fuel gas burns without leaving any residue. The gas must obviously be free of solids, but gas from coal gasification contains, apart from solids, such as coal dust and fly ash, also many other impurities, such as vaporized ash, alkali and chlorine which are detrimental to the operation of gas turbines, and it further contains gaseous sulfur compounds which are not harmful to gas turbines. All these characteristics are well-known from coal combustion, and this is the reason why very high stacks are a typical feature of coal-fired power plants.

It is therefore necessary to arrange a gas purification step between gasification and gas turbine.

2nd Condition: Gas purification ahead of gas turbine combustor.

To demonstrate the special features of the combined gasification and gas turbine process, very simple schemes based on elemental carbon as fuel have been prepared (see Fig. 1). Figure 1a shows the simplest possible gasification process. In fact, attempts have been made in the past to realize this process which consists of

G a s i f i e r

in which the coal is gasified with air under pressure.
(The product gas contains 34 % Vol CO. The gas outlet temperature is 1450°C).

G a s p u r i f i c a t i o n

In view of the prevailing temperature range, gas purification can merely effect primary removal of dust while it cannot clean the gas sufficiently to reach the purity level required for the gas turbine. This step is therefore only shown in dashed lines on the diagram.

Gasification is followed by the gas turbine process which is shown as an open-cycle gas turbine process without heat recovery, and which shall merely serve for comparison. It consists of

C o m b u s t o r

in which the gas is burnt and the maximum gas turbine inlet temperature is adjusted by the addition of air.
(An inlet temperature of 820°C can already be realized nowadays).

G a s t u r b i n e

with air compressor and generator.

W a s t e h e a t r e c o v e r y

which is necessary for economic operation of the gas turbine process.
The example shows that the exhaust gas quantity is 6 times greater compared with the combustion in conventional steam boilers. Besides, the exhaust gas

temperature at the gas turbine outlet of 394°C is still rather high. The resulting heat loss in the exhaust gas represents basically the same problem as the heat loss by condensation of the steam in steam power plants.

The simplest coal-based gas turbine process presented in the above example already demonstrates the problems which require solution for the application and improvement of the scheme.

First of all, the gas must be available under conditions which allow proper purification. The gas has to be cooled down and subjected to intensive washing whereby the solids are reduced to less than 1.5 ppm and the alkali originating from the ash is removed. This satisfies the requirements demanded by the gas turbines. Hence, the second condition has to be supplemented:

2nd Condition - Supplemental requirements:
Purification of gas by water wash
ahead of gas turbine.

This water wash is a combination of a quencher and a washer. The gas is cooled by water evaporation whereby part of the sensible heat is lost. There are several possibilities for keeping this loss at a minimum.

The most obvious possibility is to provide a waste heat boiler ahead of the washer, as it is done, for instance, in oil gasification (partial oxydation), or to transfer the sensible heat of the gas to the compressed air for the gas turbine process in a heat exchanger. However, experience with coal-fired steam boilers shows that the heat transfer surfaces tend to foul up rapidly. In conjunction with pressure gasification, the conditions are even more difficult so that this possibility can hardly be realized technically.

Another way would be to utilize the high proportion of sensible heat of the gas for endothermic gasification reactions according to the following equation:

$$C + H_2O = CO + H_2 - 52,800 \text{ BTU/lb.mol}$$

This alternative is shown in Fig. 1b. Methane formation which is involved also in pressure gasification has been neglected for reasons of simplification.

In the example of carbon gasification, the gas outlet temperature is reduced to 790 °C by the addition of steam to the gasification agent, and the loss of sensible heat on gas cooling is decreased accordingly.

Consequently, the next requirement for the gasification process is:

3rd Condition: Addition of steam to gasification agent.

To achieve the theoretical equilibrium temperature during practical operation and to ensure smooth progress of gasification at this relatively low temperature, countercurrent flow of coal and gasification agent is a necessity which can best be performed in a fixed bed. Fixed-bed gasification moreover meets the requirements regarding reaction kinetics. Another prerequisite for the gasification process therefore is:

4th Condition: Countercurrent gasification.

This requirement is even still more important when considering the gasification of coal instead of elemental carbon gasification, for the following two reasons:

Firstly, complete incineration is achieved only during countercurrent operation. Secondly, high volatile coal, for example, contains only about 65 % fixed carbon related to the d.a.f. coal, the balance being volatile matter which can be recovered by degasification during countercurrent operation and which constitutes 30 % of the heat content of the product gas.

In Fig. 1b the gas washing step is arranged downstream of carbon gasification with air and steam. Cooling down of the gas by quenching causes a loss of 15 % of the heat brought in with the fuel. This loss is only 8 - 10 % during the gasification of coal. This heat loss is not higher than in a coal-fired steam boiler, but it is still considerable.

It shall now be demonstrated that the energy loss in this gas production scheme is in reality much lower when considering the overall energy balance. To prove this, each of the two gasification processes is followed by the same simple gas turbine process.

Saturation of the gas with steam in the quencher causes an increase in the gas volume. Corresponding to this additional gas volume, the consumption of secondary air to the combustor decreases, which means that less energy is needed for air compression and that the net output of the gas turbine increases accordingly. In this connection it is further of advantage that the enthalpies and the enthalpy difference of the steam at a given temperature respectively temperature difference are greater than those of the air, so that the heat duty of the steam for a given pressure drop is higher compared with air.

A comparison between the two schemes proves this. The net output of the gas turbine is about the same in both cases. Merely the amount of sensible heat in the waste heat is less in the scheme "Gasification with steam + Gas Wash". This disadvantage can be tolerated because the exergetic value of the waste heat is relatively low, and because the complete purification of the gas effected by the wash process permits adjustment of the high gas temperature necessary for a good thermal efficiency.

Finally, the question of gas desulphurization should be considered which has been neglected so far as it is not of primary importance for the operation of the gas turbine process proper. When all other impurities have been removed from the gas, the presence of gaseous sulfur has, according to the gas turbine manufacturers, no adverse effect on the operation of the gas turbines. But the higher dew point of the fuel gas due to the presence of SO_2 and SO_3 renders waste heat recovery more difficult; this is known from conventional steam power plants.

However, the chief problem is that of air pollution. If it is possible to find technically and economically feasible solutions, this would be a real step forward.

The combination of pressure gasification with thermal power processes is a suitable way for gas desulfurization because the sulfur compounds, chiefly H_2S with a little bit of organic sulfur but no SO_2 are present in a pressurised fuel gas having an effective volume of only 1.5 % of the volume of the gas from an atmospheric combustion process. This H_2S under pressure can easily be removed by absorption in conventional wash processes and converted to marketable products, namely elemental sulfur or sulfuric acid.

The condition of the gas must of course meet the requirements for the application of the wash process for sulfur removal. These are pressures above 140 psi, temperatures of 20 - 180^{o}C and a sufficient concentration of the component to be removed. Consequently, there is another requirement for the gasification process:

5th Condition: The condition of the gas must allow the use of a wash process.

The process scheme described meets these requirements. For further particulars reference is made to Chapter 3.6. This concludes the general considerations which have demonstrated that coal gasification can be efficiently combined with a gas turbine process and that conventional methods can be used for fuel gas desulfurization. The various requirements demanded from the gas production process are summarized below:

1st Condition:	Gasification under pressure
2nd Condition:	Gas purification ahead of gas turbine by water wash
3rd Condition:	Addition of steam to gasification agent
4th Condition:	Countercurrent gasification
5th Condition:	The condition of the gas must allow the use of a wash process.

III. DESCRIPTION OF GASIFICATION PROCESS

3.1 Choice of Gasification Process

Apart from the five process requirements for coal gasification in conjunction with gas turbine power plants, which were examined in Chapter 2., there are three further requirements which concern the economics and which have to be considered when selecting the gasification process.

The investment cost for the gasification plant must not be higher than that for conventional processes which are based on the direct burning of coal.

6th Condition: The investment cost must be competitive with other processes.

Another obvious requirement is:

7th Condition: The plant must yield a profit.

Final requirement for the realization of the scheme:

8th Condition: The process must have proved its merits in practice.

A study of the available gasification processes has shown that LURGI pressure gasification meets the above requirements. A brief survey is given first on the application and technical reliability of LURGI pressure gasification, followed by a detailed description of the process.

3.2 Previous Application of LURGI pressure gasification Process

LURGI pressure gasification is a coal gasification process which has so far been applied on a commercial scale for the manufacture of town gas and synthesis gas. The process was first developed in 1933. The initial pilot plant was build in 1936 at Hirschfelde (Central Germany). This plant is still in operation for town gas production today after 33 years.

In 1938 the construction of commercial plants began. Since then, a total of 58 gasifier units for 12 plants have been built by LURGI in all parts of the world. These plants produce the following gas rates:

265 million scf/d of town gas, and
230 million scf/d of synthesis gas.

As will be seen from Fig. 2 these plants handled 40 million tons of coal up to 1969, the output of ash being 8 million tons.

The commercial plants process lignite, sub-bituminous coal and anthracite. Ash contents of up to 35 % do not create any difficulties. When producing town gas or synthesis gas, oxygen and steam are used as gasification agent.

When applied to the production of fuel gas for gas turbine power plants, air and steam can be used as gasification agent. This simplifies the arrangement and operating conditions of the plants considerably compared with town gas and synthesis gas plants.

3.3 General Outline of Scheme

A simplified flow diagram of the STEAG plant is presented in Fig. 3. The desulfurization unit is shown in dashed lines because this unit is not installed for the present due to the low sulfur content of the feed coal.

Fig. 4 shows the arrangement of the gasification plant within the power plant scheme.

The coal is gasified in the LURGI pressure gasifier with air and steam under a pressure of, say, 300 psi. The gasification pressure may be higher or lower.

The coal is fed via a lock hopper to the gasifier where it is gasified completely in countercurrent with the gasification agent. Ash is removed via an ash lock hopper. The producer gas is washed in a scrubbing cooler and saturator and is then available for the gas turbine process. The gas is saturated with steam and free of solids.

3.4 The Gasification Process

The gasification process is illustrated in Fig. 5. The gasification agent consisting of air and steam enters the gasifier through slots in the rotary grate. It flows through the ash zone arranged above the rotary grate and is then distributed over the cross-sectional area of the gasifier shaft.

It then enters the combustion zone which in the pressure gasifier is relatively narrow. Its height is about 5 times the diameter of the coal grains. This means that the residence time of the coal and the ash in the combustion zone is short.

The ash is removed continuously via the rotary grate. It is incinerated substantially completely and is cooled to about 500° C by the gasification agent. The temperature in the combustion zone is controlled by the rate of steam addition which is about 0,6 scf steam/scf air during gasification with air. In this connection it should be noted that the temperature in the combustion zone is much lower than the theoretical figure obtained from the heat balance-and under the assumption of complete combustion. The figure shows the difference between maximum temperature, which is only a fictitious value, and the real temperature. The reason for this substantial temperature difference is that, parallel to the combustion reactions, the initial gasification reactions proceed already in the combustion zone. These endothermic reactions reduce the temperature to a level of between 1000 and 1200°C, as experience has shown. Inspite of this relatively low temperature a virtually complete incineration of the ash is achieved.

The combustion gas flows upwards into the gasification zone. Its sensible heat is utilized to complete the endothermic gasification reactions which proceed according to the following equations:

1.	Boudouard reaction	$C + CO_2$	$\rightleftharpoons 2\,CO$
2.	Water gas reaction	$CO + H_2O$	$\rightleftharpoons CO_2 + H_2$
3.	Methane formation	$C + 2\,H_2$	$\rightleftharpoons CH_4$

Parallel to gasification, devolatilization of the fuel takes place. The proportion of devolatilization gas is considerable and amounts to about 30 % related to the N_2-free gas.

Depending on the reactivity of the fuel, the reactions freeze at 720 - 850°C. This is the reaction end temperature, at which a gas equilibrium is established which determines the gas composition.

Fig. 5 shows that the application of the countercurrent principle allows the utilization of the sensible heat of the gas for coal drying and preheating. Consequently, the gas outlet temperature is relatively low. It is about 500°C when processing sub-bituminous coal, and about 300° C when gasifying lignite.

At the temperature the gas leaves the gasifier. The dry crude gas has about the following composition:

CO_2	14 % vol.
CO	16 % vol.
H_2	25 % vol.
CH_4	5 % vol.
N_2	40 % vol.
	100 % vol.

This gas further contains:

> Steam from coal moisture and undecomposed steam,
>
> Tar, oil and naphtha in vaporous form,
>
> Other carbonization products of the coal, such as phenols, fatty acid, NH_3,
>
> The sulfur from the coal is present in the gas as 95 % H_2S and 5 % organic sulfur. Very little coal dust is also present.

The gasification efficiency at the gasifier outlet is about 95 %, the losses comprising 1 - 2 % losses due to unburnt matter in the ash and 3 - 4 % heat losses.

The gas is available under pressure.

3.5 Purification of Gas to Gas Turbine Purity

As the hot gas leaving the gasifier still contains little coal dust (0.01 - maximum 0.5 % wt. of the coal input) and traces of alkali and sometimes also chlorine, it must be subjected to purification treatment to make it suitable for the gas turbine process.

Pressure gasification affords complete removal of solids from the gas by quenching and washing with hot tar-containing water which is circulated. The investment cost for the required equipment is low (see Fig. 4). Cooling of the gas to saturation temperature of, say 160°C causes a loss in efficiency which can, however, be tolerated because it provides on the other hand for the gas purity which is required for undisturbed continuous operation of the gas turbine.

As higher-boiling tar fractions are condensed during cooling, the circulating wash water contains tar to which the traces of coal dust are bonded. A partial stream of the circulating water is withdrawn from the saturator and routed to a separator. The precipitated mixture of tar and dust is returned from the separator to the gasifier for cracking and gasification.

The scrubbing cooler/saturator system also removes other impurities, such as alkali and chlorines which would be detrimental to gas turbine operation. The carbonisation products from the coal, such as tar oil, naphtha, phenols, ammonia, etc. which are still present in the gas can be burnt completely. Steam saturation increases the steam proportion to 0.5 scf H_2O/scf dry gas.

Saturation with steam results in an increase in the volume of the gas. The volume of the wet gas at the saturator outlet (before entering the combustor of the gas turbine) is about 1.5 times the volume of the dry gas. While the cooling of the gas by saturation with steam causes a loss of sensible heat resulting in a reduction of the gasification efficiency, this loss is compensated in part by the increase in the gas volume which means a higher energy output from the gas turbine. This point was discussed previously.

3.6 Gas Desulfurization

The gas leaving the scrubber/saturator system is free of solids, alkali and chlorine and is suitable for the gas turbine. It still contains gaseous sulfur compounds which are not harmful to gas turbine operation but which create air pollution problems as they are emitted as SO_2/SO_3. The new and more stringent air pollution regulations require removal of the sulfur from the gas. The efforts to meet this goal in conventional steam power plants have not been successful so far, because the problems in a combustion process under atmospheric pressure are difficult for the following reasons:

a) the volume of the flue gas is relatively large,

b) the flue gas contains fly ash,

c) the flue gas is available at atmospheric pressure and temperatures of 120 - 200°C.

In the scheme using pressure gasification the problems are far less complicated and the above mentioned disadvantages are eliminated.

During pressure gasification, the coal sulfur is composed of:

90 % H_2S
5 % organic S
5 % ash sulfur

The gaseous sulfur compounds can be removed by a wash process under pressure. It is of importance that other gas consituents are not lost during the wash process. Washing with ammoniacal liquor is particulary suitable for selective removal. In this connection it is an advantage that the coal nitrogen appears in the pressure gasification gas as ammonia which means that the wash solution is a product from coal gasification. The wash system is illustrated in Fig. 6.

H_2S removal proceeds according to the following reversible reaction:

$$(1) \quad NH_3 + H_2O + H_2S \rightleftharpoons NH_4HS + H_2O$$

The wash process operates under pressure and at gas temperatures of say, 40 °C. The fat solution is regenerated by flashing and heating. The H_2S gas from the regenerator is available as feed for the Claus process to recover sulfur or for wet contact catalysis to recover directly sulfuric acid.

The presence of CO_2 in the gas renders the wash process more difficult. CO_2 is equally removed with ammoniacal liquor according to the following equations:

$$(2a) \quad CO_2 + H_2O \rightleftharpoons H_2CO_3$$

$$(2b) \quad H_2CO_3 + NH_4OH \rightleftharpoons NH_4HCO_3 + H_2O.$$

In spite of the high CO_2 partial pressure of the crude gas it is possible to remove H_2S selectively because CO_2 removal according to equations 2a) and 2b) proceeds relatively slowly. A technically feasible solution is a short residence time wash process where the gas is only in temporary contact with the wash solution. This short residence time wash process shall ensure that reaction 2a) and 2b) are incomplete while reaction 1) proceeds to the end.

The following further problem has to be considered for the realization of the proposed wash process.

The pressure gasification gas is saturated with steam at 140 - 160^{o}C which means that it contains a considerable amount of sensible heat. As low temperatures are more favourable for the wash process and for the preferential completion of reaction (1) versus reaction (2), a cooler / saturator system has been incorporated which removes the sensible heat from the gas with circulation water in a cooler and which returns the sensible heat to the gas in a saturator downstream of the H_2S removal unit.

Cost of H_2S removal

Expenditure

a)	Capital charges for the scheme presented in Fig. 6 including Claus unit (15 % depreciation and interest, 8000 h/a):	44 %
b)	Heat losses:	10 %
c)	Heat requirements for H_2S removal (calory price of fuel = 44 cents MM BTU):	24 %
d)	Electricity and cooling water:	12 %
e)	Labour + maintenance:	10 %
		100 %
	=	0.565 mills/kwh

Proceeds

a) The coal sulfur is, for example, recovered as sulfur in a Claus kiln. At 3.4 % wt. S in the daf coal and a sulfur price of $ 24,60 per sh. ton, the credit is

16.5 lb. S/MW x 1.23 UScents/lb. S = 0.202 mills/kwh

b) Credit for steam from Claus unit

$\frac{2.75 \text{ lb. steam}}{1 \text{ MW}}$ x 0.1 UScent/lb. = 0.027 mills/kwh

0.229 mills/kwh

Consequently, the cost of gas desulfurization is

0.565 mills/kwh - 0.229 mills/kwh = 0.336 mills/kwh

which has to be added to the power generation cost.

IV. APPLICATION OF NEW PROCESS USING THE LÜNEN POWER PLANT AS AN EXAMPLE

Following the description of gas production and gas purification, the process scheme of the Kellermann Power Plant of STEAG at Lünen will now be explained in detail.

4.1 Process Scheme and Design Features

The gas turbine power plant integrated with pressure gasification of coal will generate 165 megawatts at a thermal efficiency of 36 %. It shall be used to cover peak requirements for which it is well suitable because of the little time needed to start-up the gas production unit and the gas turbine. The power plant consists of the following units:

1. Gas production to handle 76 sh.tons/hr of coal with a net calorific value of 10,450 BTU/lb. and to produce 6,800,000 scf/hr. dry fuel gas.

1.1 5 LURGI pressure gasifiers - working pressure 300 psig
cross-sectional area: 30 ft^2/gasifier

1.2 Tar recycling

1.3 Gas wash

2. Expansion turbine to reduce the pressure of the producer gas from 290 to 140 psig.

2.1 1 gas heater

2.2 1 expansion turbine with compressor for the gasification air.

3. Gas turbine plant

3.1 Double combustor with

3.1.1 Gas-fired burners where the gas is burnt at an almost stoichiometric ratio.

3.1.2 Steam boiler, consisting of vaporizer and superheater. The hourly output of steam is 750,000 lb. at 1,900 psig and 525 °C.

3.1.3 Little air is added to the lower section to adjust the temperature of the combustion gases to the level permissible for the gas turbine.

3.2 Gas turbine with air compressor.

3.2.1 The gas turbine is a SIEMENS single-shaft gas turbine.

Inlet pressure		137 psig
Inlet temperature		820 °C
Output	roughly	175 MW

3.2.2 Air compressor directly coupled to the gas turbine.

3.3 Generator.

The net output of the gas turbine set is 74 MW

4. Utilization of Waste Heat.

The sensible heat contained in the combustion gas when leaving the gas turbine is utilized for two-stage preheating of the feed water whereby the temperature of the exhaust gas is reduced to 168 °C.

As the Lünen plant does not include a desulfurization step, the increase in dew point due to the SO_2/SO_3 content in the combustion gas had to be considered for waste heat utilization.

5. Steam turbine

5.1 The turbine is a condensing steam turbine with steam extraction for gasification and for feed water preheating.

5.2 Generator with an output of 98 MW.

Other typical features of the new scheme which are incorporated in the STEAG plant but which were not mentioned in the preceding chapters:

Expansion turbine and

integration of gas turbine process with steam power process.

The expansion turbine is arranged between gas production unit and combustor. The economic pressure level for gasification is above 300 psig, while with the prevailing ratio of the flow through the gas turbine to the flow through the air compressor the economic gas turbine feed pressure is about 140 psig. Consequently, the expansion of the gas in a turbine is a suitable proposition.

The incorporation of an expansion turbine is very economical in the present case because the ratio of the gas flow through the expansion turbine to the air flow through the compressor is

$$\frac{3 \text{ scf gas}}{1 \text{ scf air} + 0.5 \text{ scf steam}}$$

The output of gas volume from the gasification process is twice the quantity of the input gasification agent. Related to air only, three times as much gas is expanded as gasification air is compressed.

The increase in volume is partly due to the H_2O introduced into the gas during quenching, which was described in the preceding chapters. Moreover, the increase in volume takes place during gasification and devolatilization of the coal.

At an input of		1.0 scf gasification air
	+	0.5 scf gasification steam
		1.5 scf

the gas output is

by gasification	1.85 scf gas
by devolatilization	
(including coal drying)	0.25 scf gas
by quenching with water	0.90 scf steam
	3.00 scf

This increased volume can be utilized during pressure gasification by the incorporation of an expansion turbine whereby additional energy is provided.

This advantage which is gained from the combination of gas turbine process with gasification or reforming has been utilized in the present project only to a limited extent. The capacity of the expansion turbine could have been increased by further increasing the inlet temperature from 200 to 400°C and the pressure drop from 280 / 140 psig to 420 / 140 psig so that an extra 7 - 8 MW useful energy would be obtained. This would increase the thermal efficiency of the overall process from 36 to 37.5 %.

Another characteristic feature of the power process applied in the Lünen plant is the pressurized steam boiler. The VELOX boiler is known as boiler operating under pressure. The concept used in the present plant has, however, nothing in common with the principle of the VELOX boiler. In the VELOX boiler, increased flue gas velocities of 600 ft/sec are applied to improve the heat transfer coefficient and to thus reduce the boiler heating area. The pressure drop in a VELOX boiler is up to 45 psig.

The present scheme uses a pressurized steam boiler which operates at 140 psig gas pressure and where the ratio of pressure drop to working pressure of 0.2×10^{-2} is not higher than in normal steam boilers or heat exchangers. Initial examinations into the possibilities and economics of this pressure steam boiler were made by Prof. Drawe and Prof. Zinzen at the Technical University of Berlin in 1948. The results were very positive, but the status of technique at the time did not permit the realization of these ideas. This concept was taken up for the present project under consideration of the following major aspects:

1. By arranging a steam generator between combustor, where stoichiometric combustion takes place, and gas turbine, it is possible to remove sufficient heat from the combustion gas so that the temperature of the combustion gas can be adjusted to the level required for the gas turbine, say, 820°C. As no additional air is required, the net output of the gas turbine set increases.

2. During stoichiometric combustion, i.e. without the use of additional air for cooling the combustion gases, the waste gas rate is reduced to the minimum level possible, whereby the loss of waste gas, which is rather considerable in conventional gas turbine process, is also cut down.

3. The combination of steam power process with gas turbine process enables economic utilization of the waste heat from the gas turbine process for feed water preheating.

4. Owing to the better heat transfer coefficients, the pressure steam boiler requires a smaller heating area which makes it less costly compared with conventional steam boilers.

Hence, the integration of the thermal power process in the gas turbine process improves the efficiency and cuts down the investment costs.

4.2 Technical Data and Investment Costs

The following is a summary of the major technical data of the pressure gasification / gas turbine power plant at Lünen based on the information received from STEAG.

Technical Data of STEAG Power Plant

Coal consumption	76 sh. tons/hr
	(n.c.v. of coal 10,450 BTU/lb.)
	equal to 1,580 x 10^6 BTU/hr
Output of gas turbine	74 MW
Output of steam turbine	96 MW
	170 MW
Power required for drivers	5 MW
	165 MW
Thermal efficiency	36 %
Total heat demand	9400 BTU/kwh
Air throughput	16.5 lb./kwh
Steam consumption	4.5 lb./kwh
Combustion gas rate	18.5 lb./kwh
Cooling water consumption (gradient 8.8 $^{\circ}$C)	25.6 gal/kwh
Hence:	
Consumption of make-up water	0.465 gal/kwh
+ feed water	0.169 gal/kwh
	0.634 gal/kwh

Space Requirements

The space required for this power plant (excluding cooling tower) is only

$$175 \text{ ft} \times 160 \text{ ft} = 28,000 \text{ ft}^2$$

The built-around volumetric space is 1,340,000 ft^3.

Investment Costs

The investment costs of the Lünen power plant are 15 - 20 % lower compared with conventional power plants of same size. The capital expenditure based on the prices in 1968 and excluding gas desulfurization amounts to roughly

$ 90,--/kw

This figure includes $ 19,--/kw for gas production.

The construction period for such power plants is shorter than for conventional power plants as the gasifiers, pressure steam boilers, etc. are completely fabricated in the manufacturers' works so that erection includes only the lifting in position of the equipment and the installation of interconnecting pipework.

V. APPLICATIONS AND FUTURE DEVELOPMENTS

The example of the new STEAG power plant at Lünen is only one of the possible applications and designs. It uses process and equipment which have already proved their technical reliability on a commercial scale. It is therefore merely a first step in this new direction. Meanwhile, STEAG and LURGI have explored the possibility of how this new power plant scheme could be improved further and what other process combinations could be chosen. A brief report is given about these future developments to conclude this paper.

The efficiency of the overall scheme could be improved by introducing the following measures, amongst others:

1. In the gas production process: Increasing the capacity of the expansion turbine by preheating the gas to a higher temperature prior to expansion and by the

application of a higher gasification pressure.

2. In the gas turbine process: Increasing the gas turbine inlet temperature.

3. In the thermal power process: Application of intermediate superheating.

Recent investigations carried out by STEAG have shown that the thermal efficiency will increase to 40.5 % by intermediate superheating of the steam, and to 42-45 % by increasing the gas turbine inlet temperature.

The trend in gas turbine manufacturing towards larger scale units will permit the construction of power plants of the present type with larger unit capacities, which will reduce the specific investment costs.

The combination of gas turbine process with steam power process has been chosen because this combination improves the efficiency of the overall scheme due to the lower air compressor capacity. The same effect could be achieved by utilizing the waste heat from the turbine exhaust gas for steam saturation of combustion and gasification air.

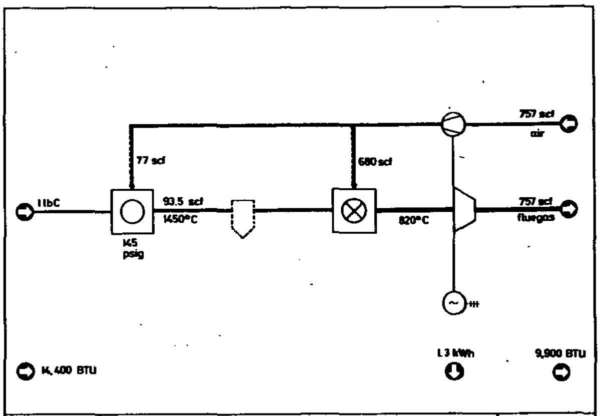
There might be cases where owing to shortage of water the combination with a steam power process cannot be realized. In the combination of pressure gasification with gas turbine process only, water consumption can be reduced to 0.1 gal/kwh which is only 10 % of the water requirements for conventional thermal power processes.

The prototype of the new power plant at Lünen shows that the combination of pressure gasification with gas turbine process requires only a minimum of time for start-up which is the reason why STEAG use this plant mainly to cover peak load requirements. As the coal can be stored and the investment costs for the power plant and in particular for gas production are relatively low, the gas-from-coal power plant can supply the peak gas load while nuclear power plants and natural gas power plants supply the basic load.

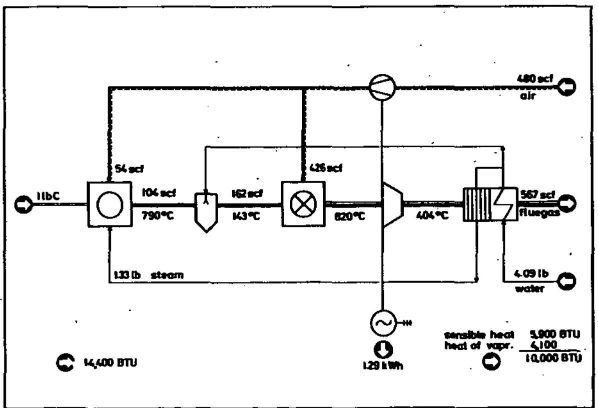
Another interesting aspect is that the gas turbine can also handle other gases, such as natural gas, coke-oven gas, etc. As the coal pressure gasification process can very well cope with load variations, the possibility of mixed operation exists.

In summing up, the following advantages of the new power plant process can be stated·

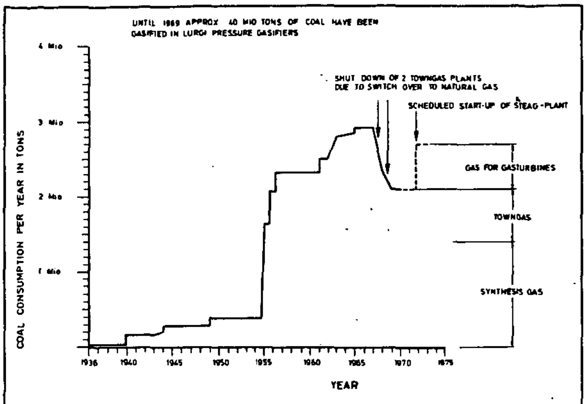
- High efficiency and low investment costs.
- Gas desulfurization can be accomplished economically so that air pollution problems are eliminated.
- It can readily cope with special conditions, such as shortage of water, peak load demand, etc.
- It offers better possibilities for further improvement than the conventional thermal power process.



(LURGI) GASIFICATION OF CARBON WITH AIR FIG. 1a



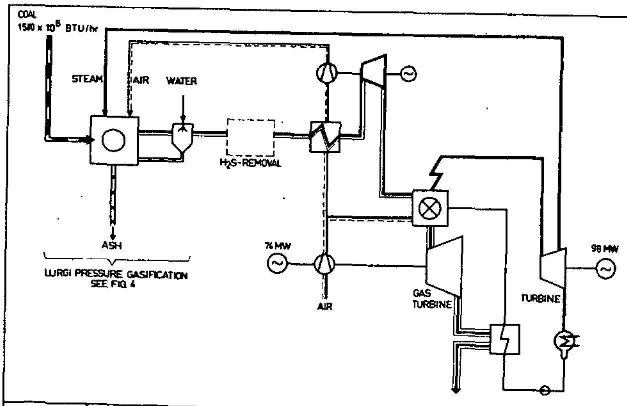
(LURGI) GASIFICATION OF CARBON WITH AIR AND STEAM FIG. 1b



LURGI

YEARLY COAL CONSUMPTION IN LURGI PRESSURE GASIFICATION PLANTS

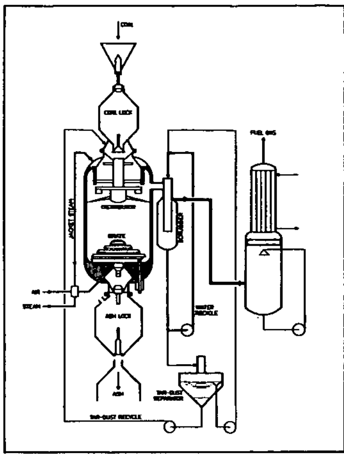
FIG. 2



LURGI

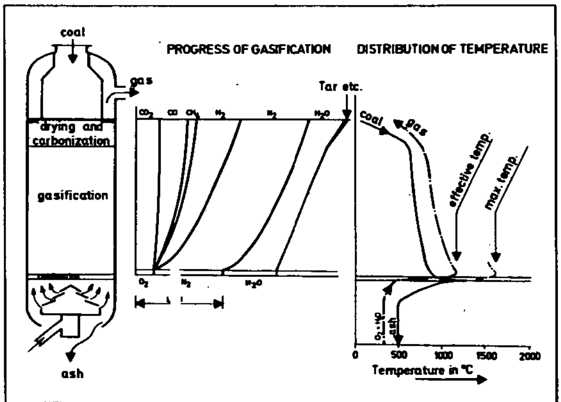
COMBINATION OF GASIFICATION WITH GAS TURBINE POWER PLANT

FIG. 3



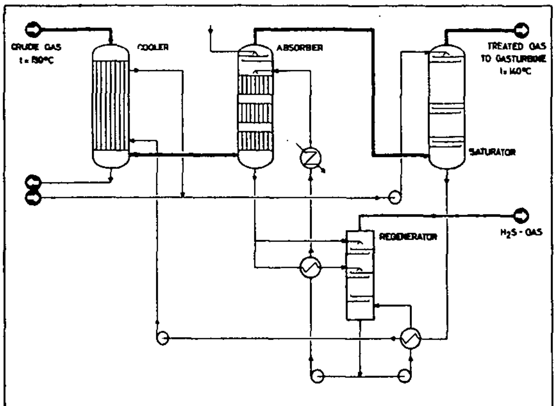
LURGI PRESSURE GASIFICATION FOR FUEL GAS PRODUCTION

FIG. 4



PERFORMANCE OF GASIFICATION IN A PRESSURE GASIFIER

FIG. 5



LURGI | H_2S - REMOVAL | FIG. 6

COMBINED STEAM TURBINE - GAS TURBINE SUPERCHARGED CYCLES EMPLOYING COAL GASIFICATION

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Introduction

Over the years, various cycles have been proposed for combining a gas turbine plant with a steam turbine plant. The major advantages advanced for such cycles were the improvement in overall cycle efficiency and the reduction in capital costs.

There are a number of combined plants in commercial operation but none are of the supercharged type and marginal economic advantages have limited general acceptance. Further, none employ coal as the single fuel.

Preliminary studies indicated that there were combined cycles which offered a better economic advantage than those then in use. Further, certain cycles appeared capable of circumventing most of the problems which had precluded the use of coal as the single fuel in existing combined plants.

In view of the tremendous potential of an attractive cycle, a program was initiated which had as its objective the development of a coal-fired, combined steam turbine-gas turbine electric generating plant with a 5 per cent lower heat rate and a 5 per cent lower capital cost than a comparable size, modern, conventional steam electric plant.

Conclusions

Sufficient development work was conducted to establish that there was an arrangement of a supercharged combined cycle which was technically feasible provided that alkali levels up to 5 ppm could be tolerated by the gas turbine.

However, during the course of the project, several important economic factors significantly affected achievement of the project's objective...

1. Coal prices decreased in many areas, thus diminishing the value of heat rate improvement.

2. Capital costs of conventional plants decreased very significantly. Thus, the potential for reducing costs below those for conventional plants was adversely affected.

Because of these factors, the economic objectives of the project could not be achieved. Consequently, further work was deferred until such time that the influencing factors changed sufficiently to alter the economic evaluation. Today, air pollution control introduces considerations which may alter the previous economic evaluation and cycles which have the potential for effective and economical air pollution control are being considered for development.

Discussion

Engineering studies had been made during a period of several years prior to initiation of this project in order to select the optimum cycle for development. Those studies concluded that a supercharged boiler cycle would afford the maximum potential for meeting the cycle efficiency and capital cost objectives. Also a specific design of gas-turbine was selected on the basis of its performance capabilities and operational compatability for integration in a large (450 MW) steam plant. This turbine was a General Electric Frame Size 8 with gas inlet conditions of approximately 1600F and 95 psia and driving a compressor delivering about 440 pounds of air per second. Two such turbines would be integrated in a 450 MW combined plant.

The usual concept of a supercharged boiler cycle in which the gas is cleaned between the boiler and the gas turbine is shown in Figure 1. In this cycle, coal is fired into a supercharged boiler where the combustion conditions, aside from the high pressure, are similar to those in a conventional boiler. All of the steam generation, superheating, and reheating is accomplished in the supercharged boiler. The hot gases from the supercharged boiler are cleaned and admitted to the gas turbine. From the viewpoint of gas turbine erosion, the gas can be cleaned adequately in a series of high efficiency mechanical collectors. However, this degree of cleaning is not sufficient to prevent gas turbine corrosion and deposits in a high temperature gas turbine. Some improvement in gas cleaning can be gained through the use of an electrostatic precipitator. However, an electrostatic precipitator will not remove volatile ash constituents which can cause deposition and corrosion.

Since cleaning of high temperature combustion gases did not appear feasible, and it did not appear possible to design a turbine intolerant to the problems of erosion, corrosion and deposition, the cycle shown in Figure 2 was considered to be more promising and was selected as the basis for development. In this cycle, all of the coal is gasified to produce a fuel gas with a heating value of about 100 BTU/SCF. The gas leaves the producer at about 900F and is cleaned with a combination of mechanical and electrostatic cleaners. The gas is then fired in a combustor, cooled to 1600F by passage over the superheater and reheater surface and is admitted to the gas turbine. The exhaust gases from the gas turbine are cooled by passage over air heater and economizer surface. Under typical conditions, feedwater is introduced into the gas producer at about 580F and leaves as steam at about 780F. The steam then flows to the combustor where it is superheated and reheated. An obvious advantage of this cycle is that the gas clean-up is performed at 900F instead of 1600F. Further, less than one-half of the gas to the turbine requires cleaning and the size of the clean-up apparatus is therefore reduced as compared to the supercharged boiler cycle. Since clean gas is fired to the combustor, the possible problems of corrosion and fouling of the superheaters and reheaters are reduced in severity.

The main problems recognized at the time that development work was outlined were those of...

1. Deriving a coal gasification process suitable for application to a power plant.

2. Corrosion in the reducing environment of the gas producer.

3. Developing a system capable of adequately cleaning the make-gas from the producer.

Initial laboratory test work on coal gasification consisted of the exploration of two gasification processes. These are shown in Figure 3. The first of these was a fixed-bed process in which coal was fired, with theoretical air, in a lower furnace. The hot gases passed upwards and countercurrently to a coal bed, several feet thick, and fed from above. The coal bed was supported on a water-cooled tubular grate and was operated with the lower portion slagging.

The second gasification process was a suspension system which also utilized a lower furnace in which fuel was fired with theoretical air. The hot combustion gases passed upwards and crushed coal of the sizing of 1/4 inch x 0 was injected into these gases at the outlet of the primary furnace. The gas velocities were maintained sufficiently high to keep the coal in suspension. This gasifier was constructed with an annular space between a silicon carbide tube, in which gasification took place, and an outer jacket. Combustion gases from a natural gas burner at the top flowed through the annulus to reduce heat losses from the gasification zone.

The experimental results obtained from the operation of these two gasifiers revealed that the fixed-bed process provided a somewhat richer gas than that obtained from the suspension process. However, the fixed-bed process produced tars which were considered troublesome whereas the gas from the suspension process was tar-free. The processes also were evaluated on the basis of their suitability to large power plant application and from the standpoint of adaptability to a wider range of coal properties and coal sizing and considerations of design, construction and operation, the suspension process was selected as the better choice.

The next step in gas producer development consisted of the design and construction of a large suspension gasifier with a coal gasification rating of about 5000 pounds per hour. This gasifier went into operation on June 1961 and an isometric view of the apparatus is shown in Figure 4.

The main components of this apparatus were the gasifier in which the make-gas was produced and the combustor, in which the make-gas was burned. Air from the forced draft fan passed through the primary section of the air heater and a portion was supplied as combustion air to the combustor. The remaining portion passed through the secondary section of the air heater and was supplied to the gasifier at temperatures up to 1000F. The coal was pulverized in an air swept mill and conveyed with primary air to the burners. The gas produced was cooled over convection heat absorbing surface to about 800F and entered two 42 inch cyclone collectors where the coarse fly-coke was removed from the gas. The fly-coke was collected in a hopper, fed through a rotary feeder and reinjected into the gasifier. The make-gas leaving the cyclone collectors was conducted to the combustor where it was burned with excess air.

This equipment was operated for about two years during which time a number of configurations of the gasifier were explored. The original arrangement consisted of a horizontal Cyclone Furnace firing into the gas producer shaft. All of the coal was injected into the base of the shaft and the fly-coke, which was separated from the make-gas, was refired into the Cyclone Furnace. Since the coal consumption in this gasifier was about 5000 pounds per hour, a single Cyclone Furnace was selected to avoid the combustion problems with multiple smaller-sized Cyclone Furnaces. However, the single Cyclone Furnace arrangement introduced gas flow distribution problems which would not exist to the same degree with multiple Cyclone Furnaces. Consequently, the final arrangement, Figure 5, with the horizontal Cyclone Furnace included a transition section between the Cyclone Furnace and the secondary furnace so constructed as to convert the gas spin on the horizontal axis into a gas swirl on the vertical axis.

Gas producer theory shows the very strong effect of gasification zone heat losses upon the heating value of the gas produced; and analysis of the horizontal Cyclone Furnace gasifier arrangement indicated that lower heat losses might be expected by using a vertical Cyclone Furnace firing upward into the gasification shaft.

At the same time, considerations based on theory and practice resulted in the recognition that the vertical Cyclone Furnace would have to operate at a lower rating than the horizontal Cyclone Furnace and that finer coal sizing would be required in order to prevent undue carbon loss to the slag. On the basis of these analytical studies and information obtained from plant visits and surveys of the operation and performance of modern European gas producers, it was decided to convert the horizontal Cyclone Furnace type producer to the vertical Cyclone Furnace type in order to explore the possible advantages of this arrangement and the configuration is shown in Figure 6.

The operation of this producer did not show any striking difference in performance. Both producers operated with acceptable carbon loss to the slag and the range of gas heating values obtained were comparable and of the order of 70-80 BTU/SCF. Extrapolation of these results to the lower percentage heat losses in a gasifier of commercial size predicted that gas with a heating value of 100 BTU/SCF could be expected from either type. The vertical Cyclone Furnace produced somewhat less lamp black but this, in itself, would not dictate the choice between the two. The choice involves consideration of other factors, foremost of which are the comparative costs and the producers, the associated fuel handling systems and the simplicity of operation. Summing up the results of the gas producer development work, two alternate types of gas producers were developed, either of which is applicable for use in a combined steam-gas turbine cycle of commercial size.

Investigations into the problem of corrosion in the reducing atmosphere of the gas producer consisted first of a literature search. Because of the difference in metal temperatures and partial pressures of the gas constituents, almost no previous gas producer corrosion experience could be found which applied under the conditions expected in a gas producer for a combined cycle. However, some petroleum refinery experience at the temperatures and hydrogen sulfide concentrations which were expected was available. The corrosion rates reported from carbon steel, the intermediate croloys, and even for the common austenitic stainless steels were discouraging. However, though the refinery experiences were at the hydrogen sulfide levels which were expected, the partial pressures of the other gas constituents were much different from the expected conditions. Experiments were therefore designed to test various alloys under conditions duplicating those expected in a commercial producer. The tests were conducted in autoclaves under the conditions of pressure, temperature, and gas composition expected in the commercial producer. These tests substantially confirmed the reported refinery experiences. Search for better alloys in subsequent tests ultimately led to two alloys which exhibited satisfactory corrosion resistance. The first of these was an 18 CR - 13 Ni steel with 2.5 per cent silicon. The second was an 18 CR steel with 4 per cent aluminum. These steels exhibited corrosion rates of about 0.003 inches per year at 950F metal temperature in the atmosphere expected in a gas producer fired with a 5 per cent sulfur coal.

The third major area in which development work was undertaken was the clean-up of the make-gas from the producer. The original concept for cleaning the make-gas to the degree required for the series gasifier and combustor cycle described earlier involved the combination of mechanical collectors followed by an electrostatic precipitator. It was recognized that the electrostatic precipitator involved the major difficulties expected. Therefore, an electrostatic precipitator was designed and built to investigate cleaning of the gas from the producer. Problems were immediately encountered in the way of insulator electrical shorting due to deposits of carbon black. This difficulty was largely overcome by employing a charged grid around the insulator together with gas sweeping using nitrogen as the purge gas. A small number of performance tests were conducted on the precipitator and the results indicated that the permissible gas velocities were so low as to make the precipitator for a commercial unit very large and prohibitively expensive.

It then was decided to determine whether the gas clean-up could be accomplished to a sufficient degree by mechanical means alone. Test apparatus was installed to determine the effectiveness of mechanical cleaning of the make-gas from the standpoint of turbine erosion. The apparatus, as shown in Figure 7, consisted of a series of mechanical collectors, a combustor where the producer gas was burned, a heat exchanger to cool the gas to the desired temperature entering the grids, a turbine grid simulating the first stage nozzles and blades and a steam ejector to produce the desired gas velocities through the grid. Test results indicated that the make-gas could be cleaned by mechanical means alone to the degree required to prevent gas turbine erosion.

However, it was recognized that cleaning of the make-gas by mechanical means only could introduce serious problems in the cycle originally selected for development. Two possible problems which were envisioned were...

1. turbine erosion due to ash agglomeration and subsequent spalling of coarse particles from the combustor convection surfaces and

2. corrosion in the gas turbine due to the build-up of alkali in the system.

The cycle, shown in Figure 8, was conceived to circumvent these difficulties. This cycle can be described as a parallel gas producer and supercharged boiler arrangement. In this cycle, the major portion of the coal is consumed in the supercharged boiler under normal conditions of excess air. The combustion gases are then cooled to 900F and cleaned in an electrostatic precipitator. Since the fly ash is free of carbon, the operation of this precipitator does not present the problems encountered when cleaning gas from the gasifier. In addition, the gas temperature is sufficiently low that volatile ash constituents are essentially absent and the alkali can be collected as a fume and discharged from the system. The operating temperature of the precipitator would not present difficulties due to electrical characteristics of the gas or ash.

Sufficient coal is gasified in the gas producer to supply the combustor with enough fuel to reheat all of the gas to the turbine to the desired inlet temperature. The gas turbine exhaust gases are cooled to the stack temperature with air heater and economizer surface in a manner similar to the series cycle.

Under typical conditions, feedwater enters the gasifier at 580F and leaves at 670F. It then passes to the supercharged boiler where the superheating and reheating takes place.

The parallel cycle possesses a number of important advantages over the series cycle. Perhaps the chief one is the simplification in the gas cleaning. In the case of the parallel cycle, alkali is rejected from the cycle along with the fly ash from the precipitator in addition to its disposal with the slag. The curves of Figure 9 show the relationship between the alkali concentration to the gas turbine and the make-gas cleaning efficiency for the parallel cycle with the assumptions indicated. The assumptions require a 95 per cent efficient mechanical collector to reduce the alkali to the turbine to 5 ppm when burning a 0.25 per cent total alkali coal. Further testing under gas turbine conditions of pressure and temperature would be required to assess whether an alkali level of 5 ppm in the gas to the turbine could be tolerated.

Since the gas producer requires stainless steel to provide corrosion resistance, it is a costly component in the cycle. In the parallel cycle, about 30 per cent of the coal must be gasified as compared with the need for 100 per cent gasification in the series cycle. The size and cost of this component, therefore, are reduced in the parallel cycle. This advantage is further augmented by the reduced temperature pick-up in the gasifier cooling circuit. The resulting lower metal temperature limits the corrosion rate to a tolerable level.

The parallel cycle presents further advantage in the way of the increased operating flexibility possible in the gasifier. Since the fly-coke removed from the make-gas is fired to the supercharged boiler, the gas producer need not operate under the condition of 100 per cent carbon utilization. This permits operation with a higher fuel to air ratio in the gasifier which produces gas with an increased heating value. In effect, the gas producer can be operated anywhere between the conditions of a gas producer or a carbonizer.

Evaluation of the considerable data obtained as a result of the research and development led to the assessment that a large scale plant, of the parallel cycle type, would be technically feasible provided that alkali levels up to 5 ppm could be tolerated by the gas turbine.

Engineering designs and studies closely paralleled the laboratory work throughout the entire development and analysis of the parallel cycle showed that the desired heat rate reduction could be obtained.

To determine whether the commercial development of this cycle could be justified, a 450 MW plant was designed to a sufficient degree that reliable cost estimates and evaluations could be made. Substantial engineering effort was expended in the design of all the plant components to assure functional and structural adequacy.

Sketches of the side elevation and the plan view of the plant arrangements which were developed are shown in Figures 10 and 11 respectively, and an artist's sketch of the plant is shown in Figure 12.

From these studies it was concluded that...

1. The machinery arrangement for a combined plant involves more components, is more complex and is inherently more expensive than that of a conventional plant.

2. A combined plant does not offer a substantial saving in the cost of plant components external to the boiler plant and steam generator.

3. The increased cost of the plant was greater than the value of the heat rate improvement.

4. The reductions made in the cost of conventional plants during the course of this development significantly affected the cost comparison between conventional and combined cycles.

5. The significant decrease in the average cost of coal delivered to utilities which occurred during the course of the project decreased the worth of heat rate improvement and was unfavorable to the combined cycle economic comparison.

A very thorough analysis of the economic and market evaluations concluded that the cycle did not offer sufficient economic inducement to justify the very large expenditure that would be required to continue the development to reach the commercial product stage. Accordingly, it was agreed that development should be discontinued until such time that major factors altered sufficiently to change the above conclusion.

The increasing emphasis on the control of air pollution has resulted in renewal of interest in combined cycles of the supercharged type which offer the potential for removal of the pollutants from gases at elevated pressures and of reduced volumes.

There are a number of cycles which have been proposed for this purpose and an example of one is shown in Figure 13. In this cycle, coal and air are fed to a pressurized, water-cooled gas producer which delivers combustible gas at about 900F. The particulate matter is then removed either mechanically or by filtering if filter media capable of operating at this temperature are developed. The sulfur compounds can be removed by solid adsorbents of the metal oxide type which can be regenerated to produce sulfur dioxide suitable for feed to a sulfuric acid plant. Alternately, regeneration to form elemental sulfur may be feasible and this is under investigation.

The clean combustible gas is fired in a combustor which discharges to a high temperature gas turbine which exhausts to the steam generating and heat recovery portion of the system. The water and the steam side of the cycle have been omitted from the figure for the sake of simplicity. Since the flame temperature in the combustor is less than that in a conventional coal-fired boiler furnace, significantly less nitrogen oxides will be produced.

Since the gas is produced and cleaned at high pressure, the size and possibly the cost of the gas producer and cleanup system would be significantly less than with atmospheric pressure systems. Further, the power cycle is more efficient and this, coupled with the value of sulfur recovered, indicates promise for an economical solution to the air pollution problem of the electric utilities.

Many of the important areas of this system have been developed through the pilot plant stage and the cycle is considered to be technologically feasible. However, the economic evaluation of the cycle and the development costs which would be required have not been examined in sufficient detail to permit conclusions concerning the commercial potential. Perhaps, with the application of sufficient engineering ingenuity, a cycle of this or a similar type may become the economical power plant of the future.

Acknowledgment

The development program which provided the basis for this paper was jointly sponsored by Babcock & Wilcox and General Electric. The efforts and technological contributions of the joint project team, consisting of members from both companies, are gratefully and respectfully acknowledged.

COMBINED STEAM-GAS TURBINE CYCLE WITH COAL-FIRED SUPERCHARGED BOILER

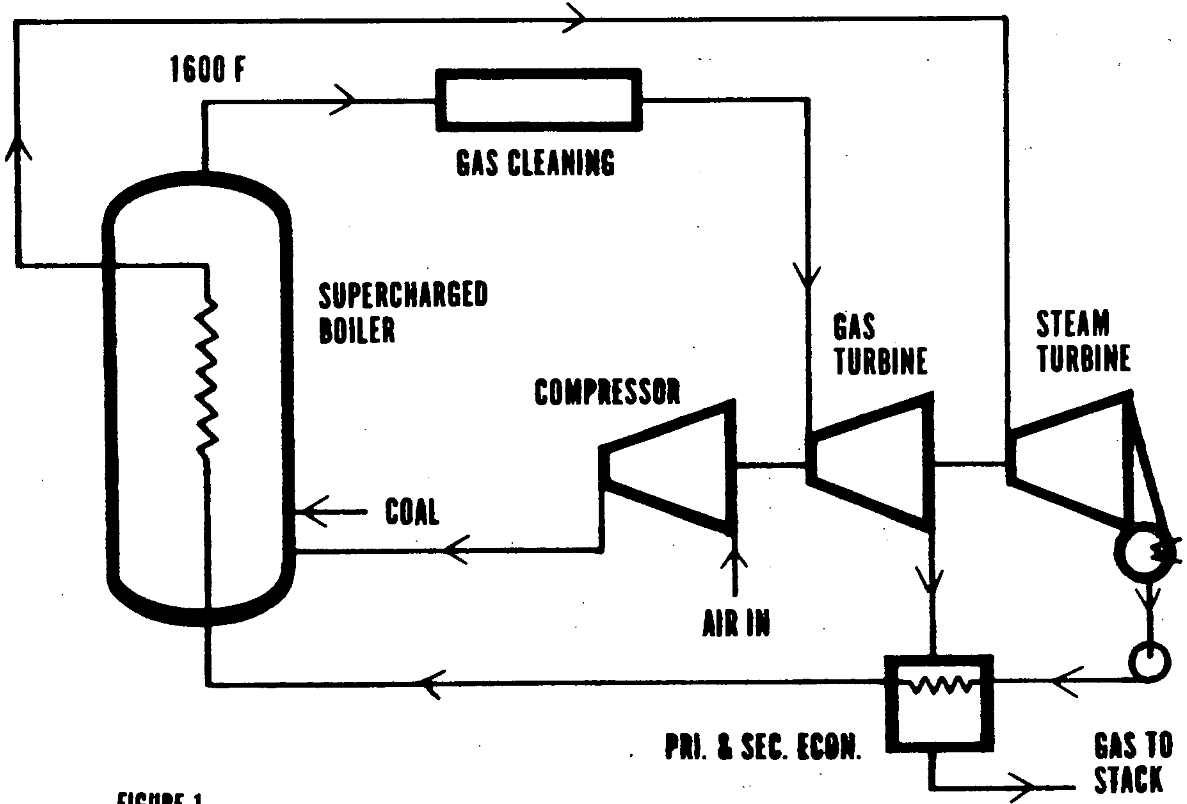


FIGURE 1

SUPERCHARGED GASIFIER COMBINED STEAM-GAS TURBINE CYCLE

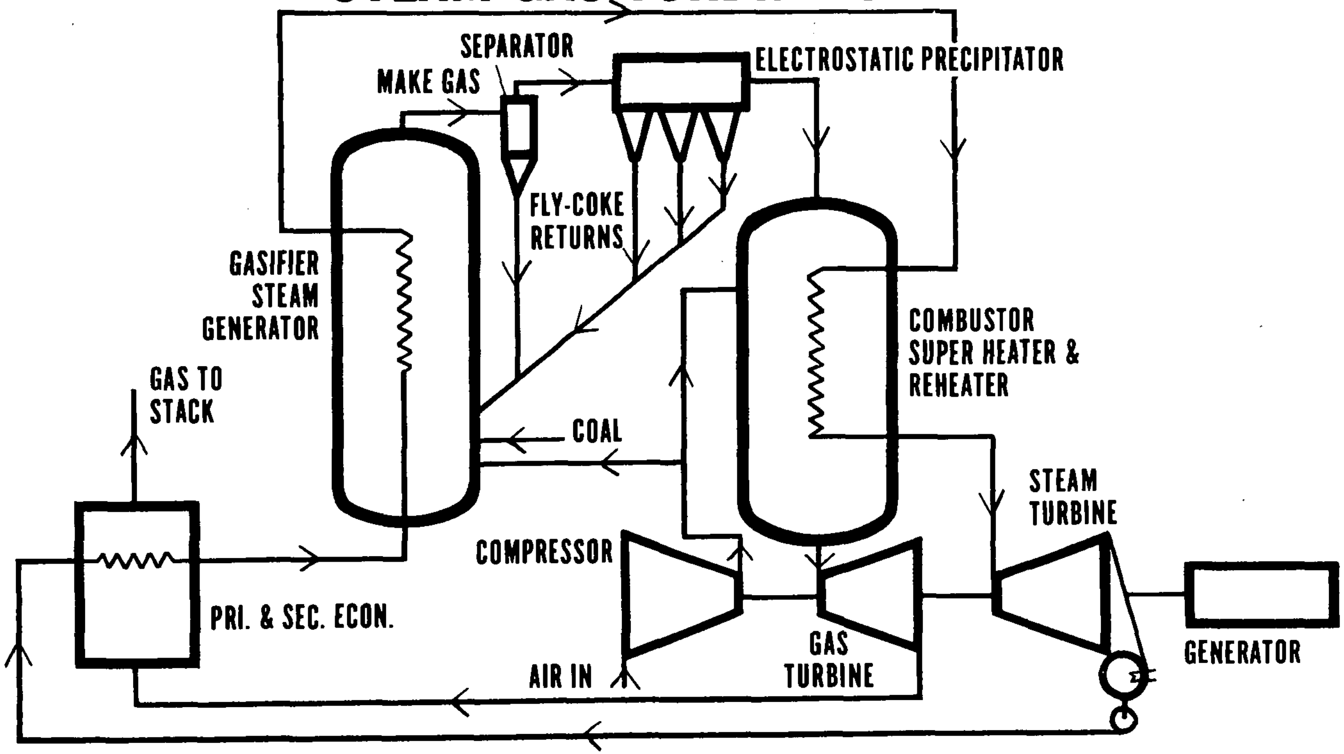
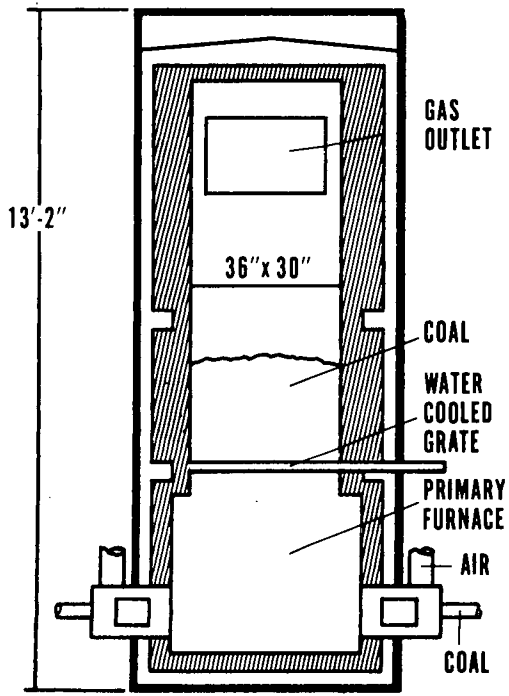


FIGURE 2

FIXED BED GASIFIER



SUSPENSION GASIFIER

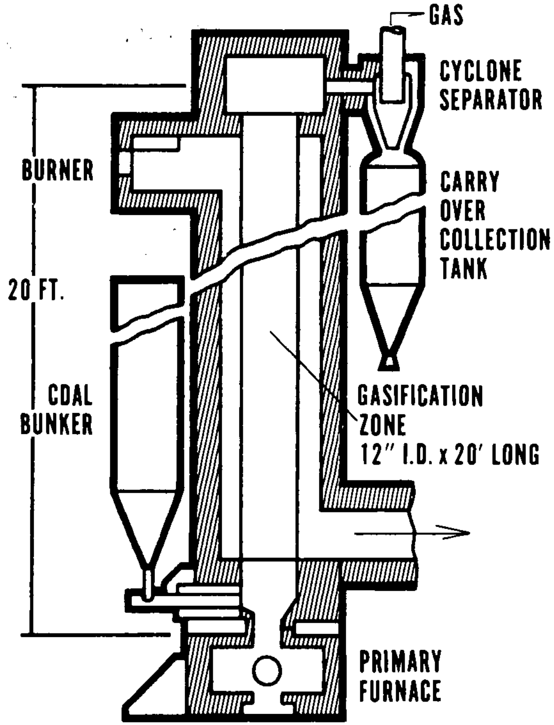


FIGURE 3

ALLIANCE LARGE SCALE GASIFIER

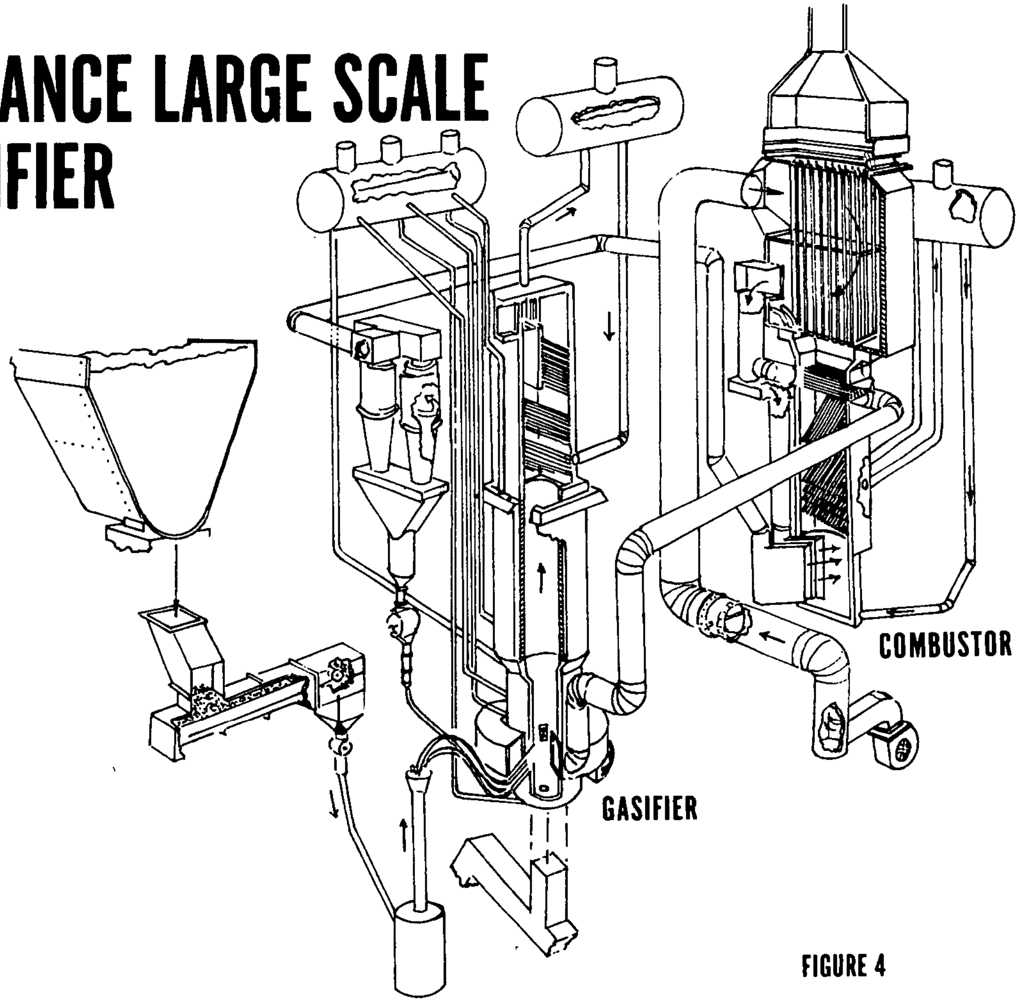


FIGURE 4

VERTICAL VORTEX GASIFIER

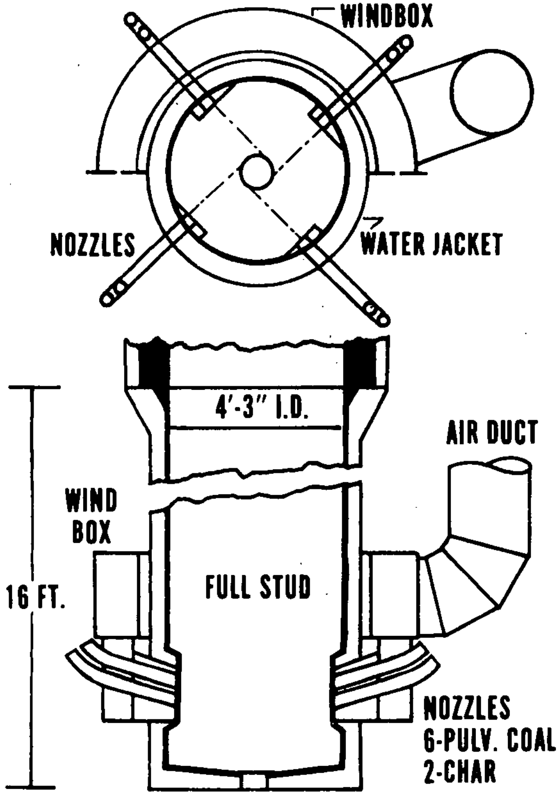


FIGURE 6

HORIZONTAL CYCLONE GASIFIER

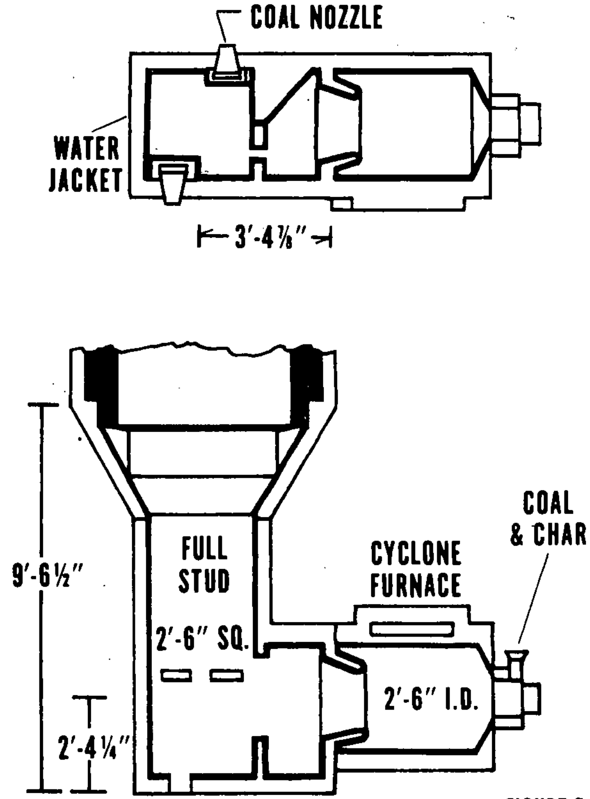


FIGURE 5

GAS CLEANING AND TURBINE GRID TEST

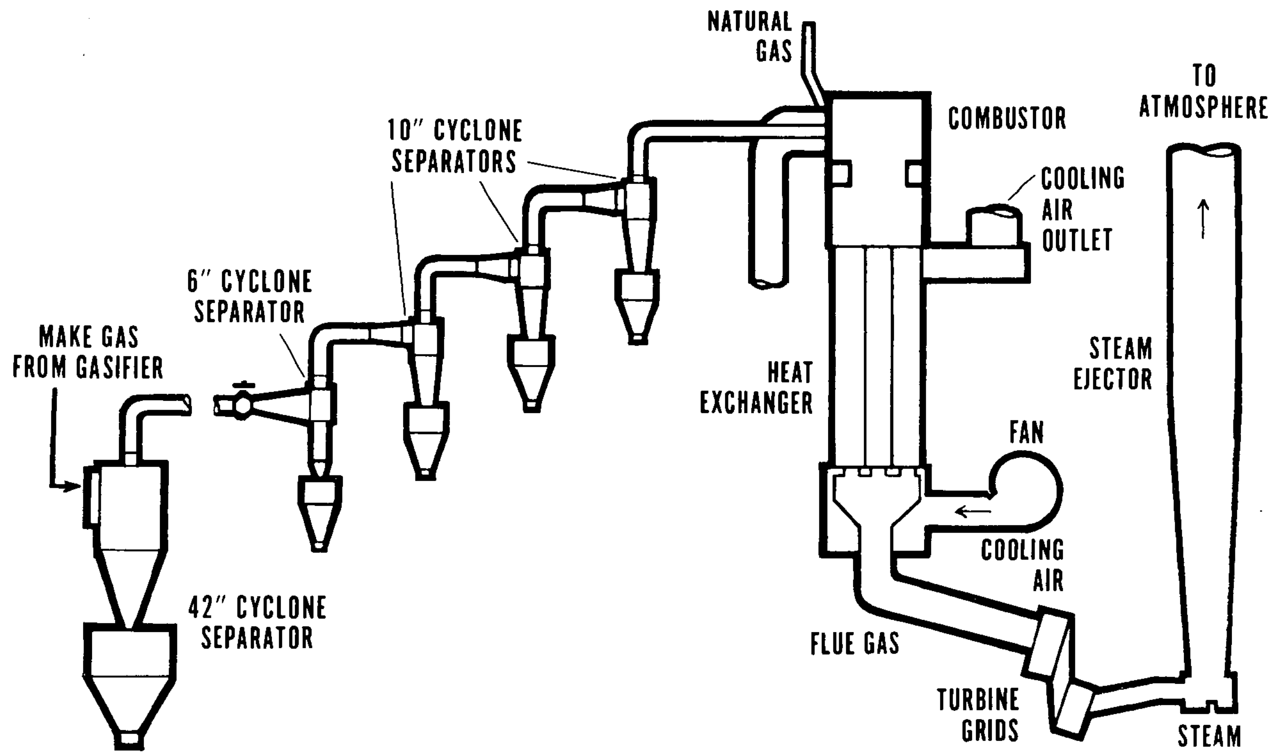


FIGURE 7

COMBINED STEAM-GAS TURBINE-CYCLE, WITH SUPERCHARGED GASIFIER AND BOILER

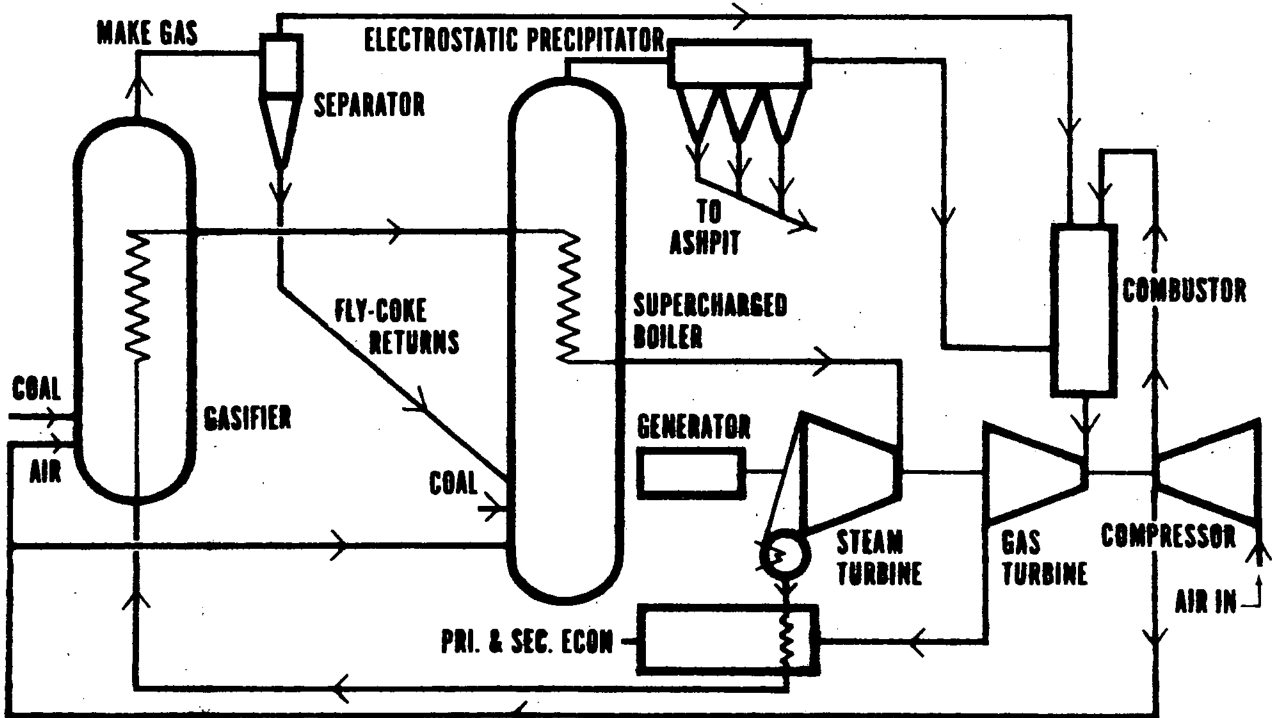
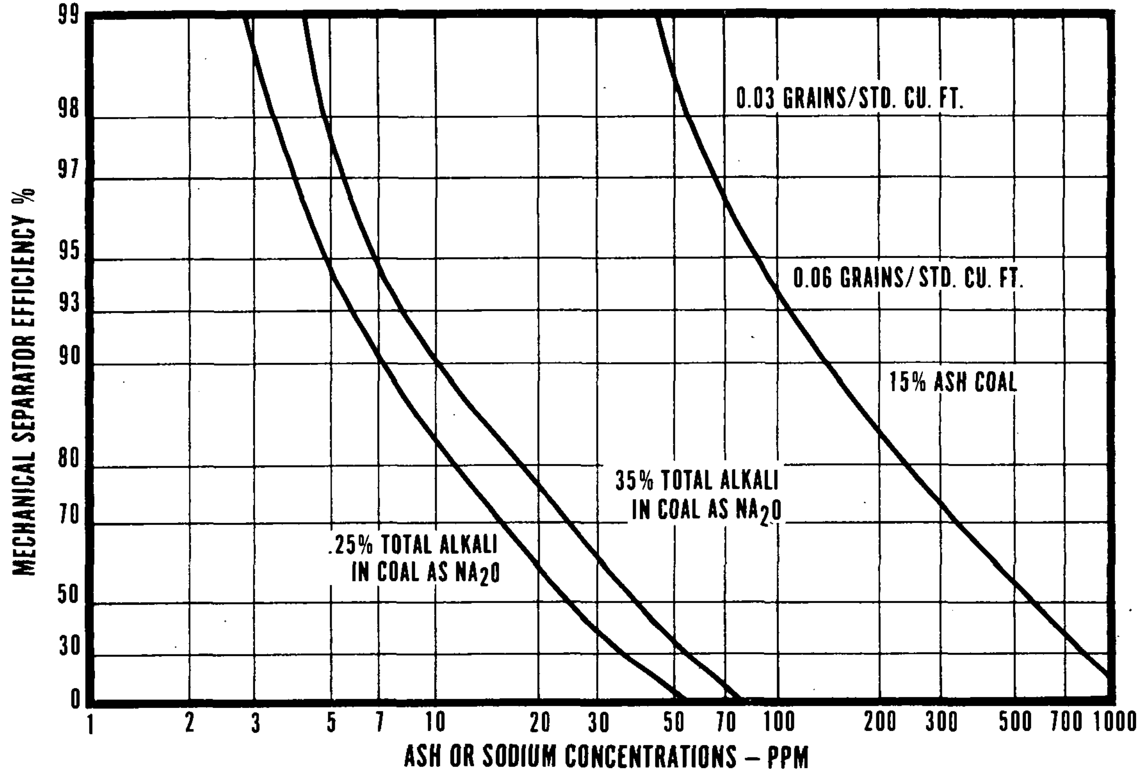


FIGURE 8

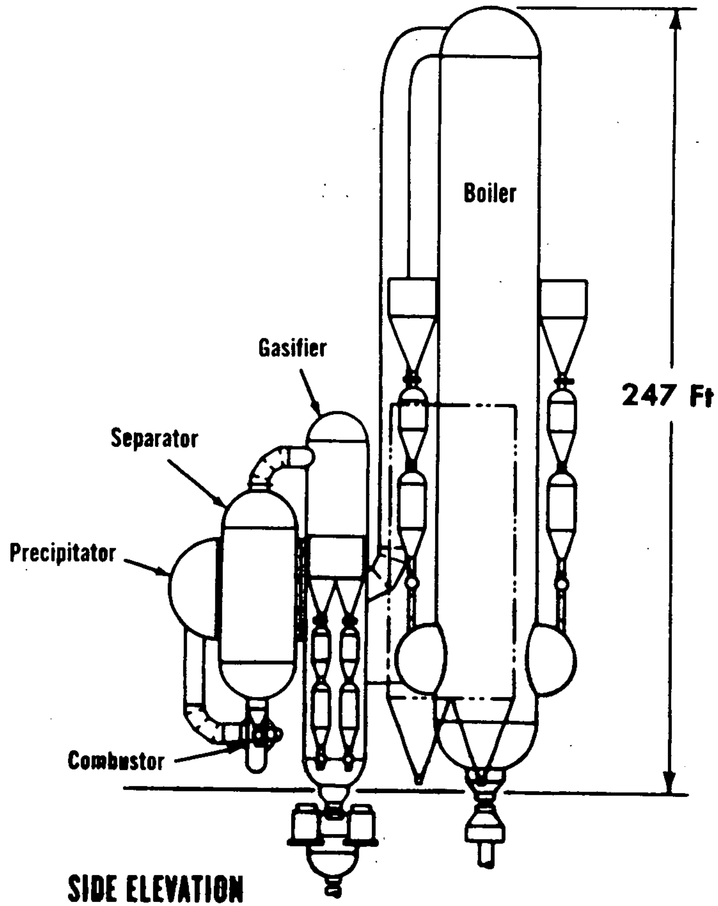
CONCENTRATIONS TO GAS TURBINE



ASSUMED CONDITIONS

ELECTROSTATIC PRECIPITATOR EFF. 90%
GASIFIER ASH SLAGGING EFF. 80%
SUPERCHARGED BOILER ASH SLAGGING EFF. 85%
SUPERCHARGED BOILER & GASIFIER ALKALI SLAGGING EFF. 25%

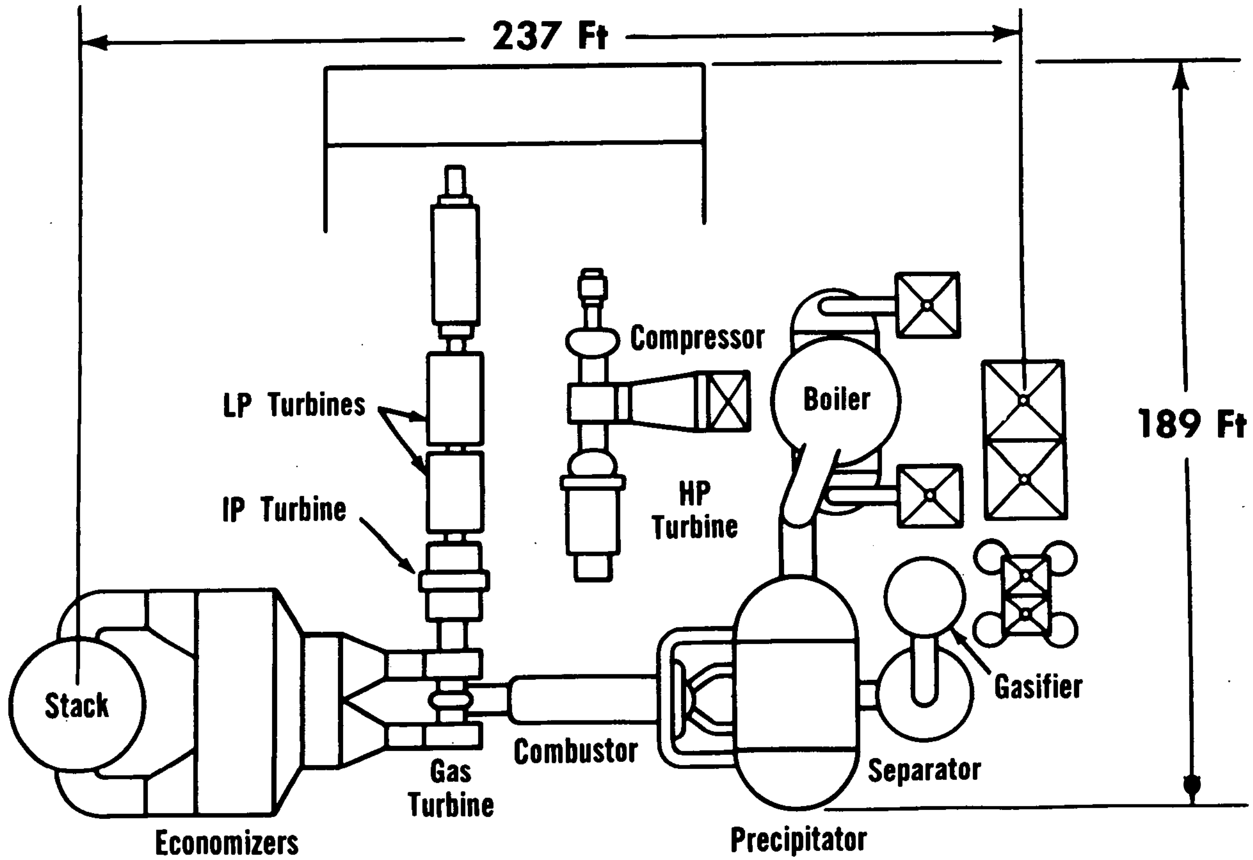
FIGURE 9



SIDE ELEVATION

450 MW STEAM-GAS TURBINE PLANT

FIGURE 10



Plan View

450 MW STEAM-GAS TURBINE PLANT

FIGURE 11

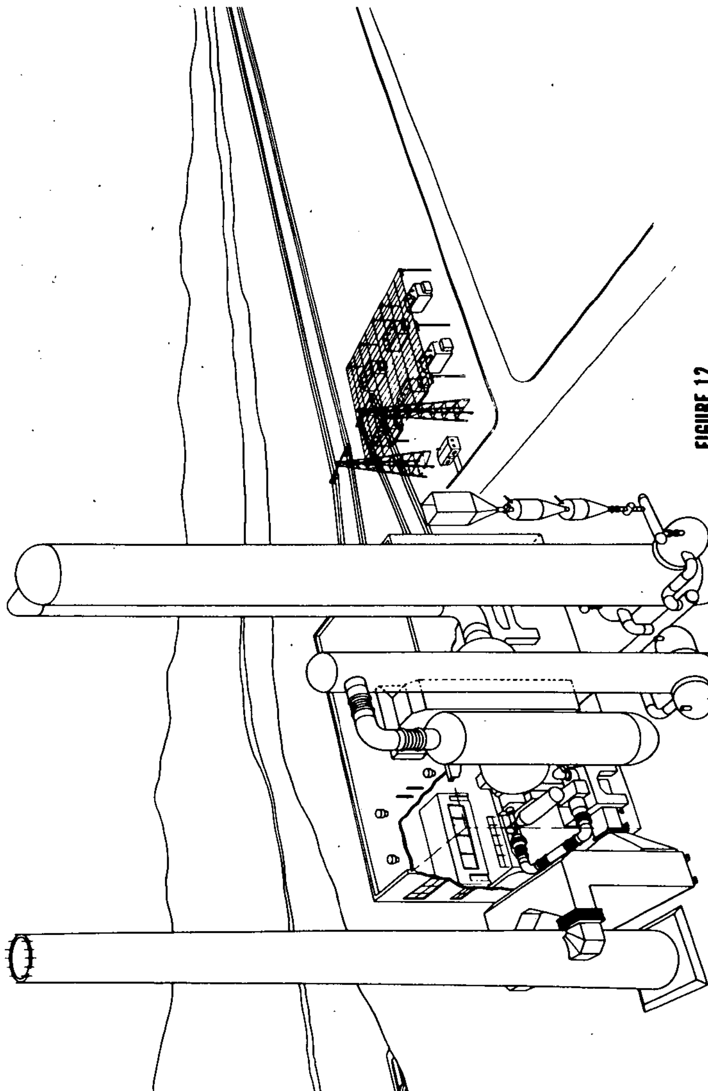


FIGURE 12

450 MW STEAM-GAS TURBINE PLANT

ADVANCED POWER CYCLE FOR AIR POLLUTION CONTROL

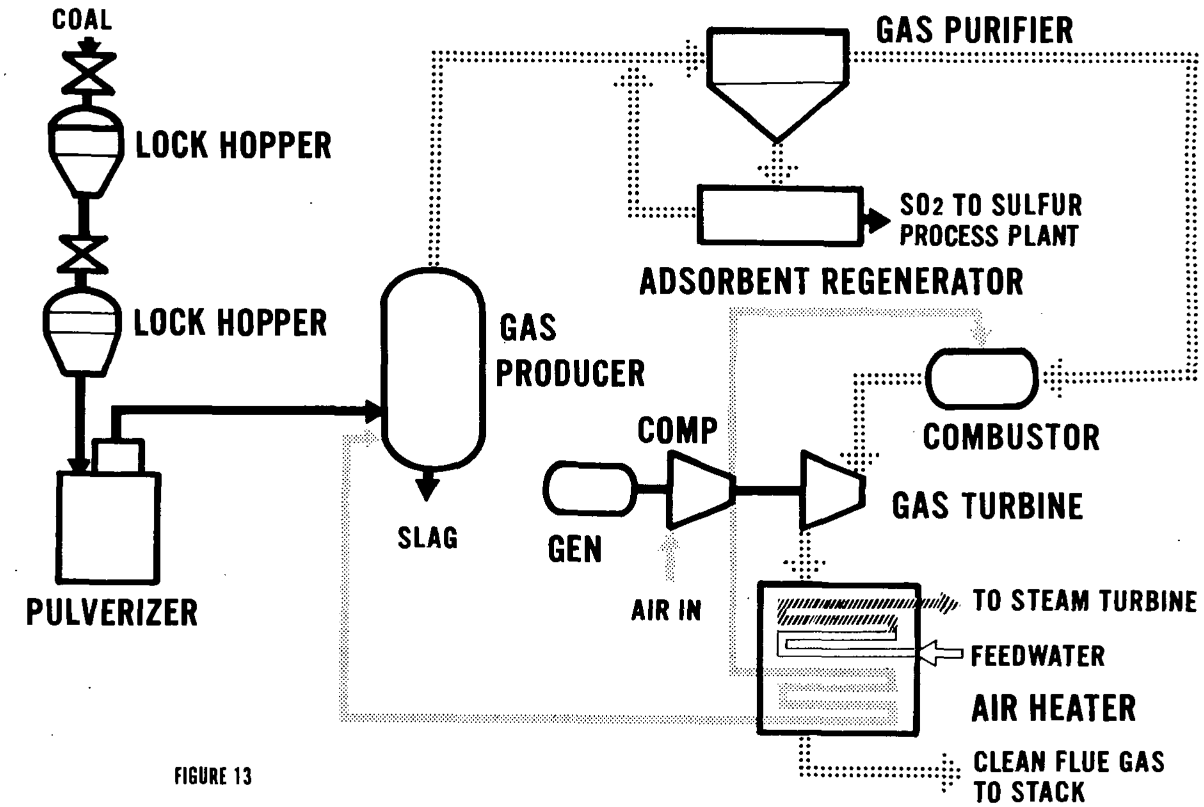


FIGURE 13

FLUIDISED COMBUSTION UNDER PRESSURE

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1. INTRODUCTION

Over the last two decades, fluidisation has become a widely accepted and reasonably well-understood means for bringing about mass and heat transfer in the chemical and petroleum processing industries.

Where uniform temperature, good mixing, high heat transfer rates, large areas of reaction surface, and mobility of reacting species are important, it would be difficult to envisage a more suitable processing tool. It cannot, of course, be inferred that such systems do not have many problems, or that all the potential benefits of fluidised bed processing systems have been successfully realised. The literatures contain many dissertations, (e.g. Squires, 1961) written by those who have encountered difficulties in coming to terms with fluid beds.

The purpose of this paper is to outline the potential benefits of fluidised combustion, mainly in the field of power generation, and particularly when it is carried out under pressure. We also indicate some of the problems that have to be overcome to realise these benefits, and outline some of the investigations being made to solve them.

2. FEATURES OF FLUIDISED COMBUSTION

The fluidised bed for a combustion system can be formed from any inorganic particulate matter, e.g. mineral matter from coal, crushed refractory, limestone or dolomite. The particles should preferably have good resistance to attrition, otherwise excessive quantities of material would need to be fed to make up for elutriation of fines and to maintain bed level.

Coal (or oil) injected into the fluid bed is burned at high rates of heat release and the rates of heat transfer to surfaces immersed in the bed are much higher than the average rates attained in conventional furnaces.

Fluidised combustion is not an entirely new technique. Over the last thirteen years, fluidised combustion systems have been described for burning difficult fuels such as anthracite fines (Stouff, 1957), lignite (Panoiu and Cazacu, 1962; Novotny, 1963) and washery tailings (Fassotte, 1961). Plants described by Godel (1963) and by Okaniwa and Suzuki (1959), which have apparently proved successful, differed from the present concept in that the combustion temperatures were high enough to fuse the ash, and all the heat release was recovered from the gas leaving the bed, using more or less conventional boiler heat exchange surface. Another successful plant (von Friese, 1961) contained cooling tubes in the bed to reduce clinkering, and about half the total steam output was generated in these tubes.

In the present concept of fluidised combustion, most of the heat is extracted from the bed in this way. Though foreshadowed to some extent by the little-publicised activities of the Badische Anilin und Soda Fabrik (1957) and of Combustion Engineering Inc. (1957), this concept largely stems from investigations started about six years ago in the U.K. at the research laboratories of the Central Electricity Generating Board (Botterill and Elliott 1964), and continued at the Central Research Establishment (CRE) of the National Coal Board and at the British Coal Utilisation Research Association (BCURA). The work at BCURA was originally aimed at developing small packaged industrial boilers (Parker, Roberts and Wright, 1969), whereas the other activities have been concerned with fluidised combustion of coal for large power plant (McLaren and Williams, 1969). For the last two years, BCURA have also been investigating fluidised combustion under pressure, primarily for power production (Hoy and Roberts, 1969).

Fluidised combustion of oil has been pioneered in the U.K. by Esso (Moss, 1968). The system proposed, usually known as the "Chemically Active Fluidised Bed Combustor", involves burning residual oil in a fluidised bed of limestone or dolomite, which absorbs sulphur dioxide. The lime is regenerated, with recovery of the sulphur, by treatment in a separate stage. A similar arrangement has been proposed for burning finely pulverised coal.

2.1. Potential Advantages of Fluidised Combustion

The present concept of fluidised combustion has many advantages over conventional combustion systems. By operating fluidised combustors under high pressure, many additional benefits can be gained. The most important of these are the reduction of plant size and the possibilities for improving thermal efficiency and simplifying some features of fluid bed operation.

All the potential benefits accrue from the following key factors:

a. The large surface available for reaction, long solids residence time, and excellent solids mixing, enable high combustion efficiency and intensity to be achieved at combustion temperatures as low as 1300-1800^{o}F (700-1000^{o}C).

b. The fluid bed provides high rates of heat transfer to immersed surfaces.

<u>Combustion Intensity and Efficiency</u> - The combustion intensity (rate of heat release per unit bed volume) is proportional to the mass velocity of combustion air, and therefore to the fluidising velocity and operating pressure.

Fluidisation velocity depends mainly on the particle size distribution of the material that forms the bed. Unless inorganic material in addition to that associated with the coal is supplied to the bed, the size consist of the bed material will depend upon the size of the ash associated with the coal and its resistance to attrition. Fluidised combustion in its simplest form therefore is better suited for burning uncleaned coals than for burning low-ash clean coals.

The operating range is from the lowest velocity needed to maintain fluidisation to the velocity above which elutriation is excessive. Though the lower limit is relatively insensitive to pressure (Fig. 1), increasing the pressure can increase the quantity of material elutriated from the bed (Fig. 2). With operation under pressure, therefore, rather more carryover material has to be recycled to attain the same level of combustion efficiency.

The fluidisation velocities and heat release rates relevant to the three size ranges of coal that have been used in most of the U.K. investigations are as follows :

Nominal coal size range	in.	1/16 - 0	1/8 - 0	1/4 - 0
Fluidising Velocity range	ft/s	1 - 4	2 - 9	5 - 12
Heat release rate	10^6Btu/ft^{2}h	0.09-0.36	0.18-0.81	0.45-1.08

(Bed temp 800°C; 10% excess air)

Combustion intensity depends upon two further factors, (i) the minimum depth of bed needed to get good dispersion of the fuel and to complete release and combustion of volatile matter; this is the main factor at atmospheric pressure, and (ii) the depth of bed required to accommodate the heat transfer surface needed for attaining the chosen bed temperature; this is the main limiting factor at high pressures.

At atmospheric pressure combustion efficiencies in excess of 99% have been obtained when burning 1/16"-0 coal in beds about 2 ft deep (McLaren and Williams). Combustion intensities in excess of 0.1 million Btu/ft^{3}h, some ten times those allowable in conventional boiler plant, are achieved. Burning 1/4"-0 coal the combustion intensity can be at least five times higher, albeit at some sacrifice of combustion efficiency.

At high pressures the potential reduction in boiler size is dramatic (Fig. 3), a fluidised bed boiler operated at 15-20 atm could be about 1/15th the volume of a conventional boiler operating at atmospheric pressure.

The size of the boiler 'envelope', however, is not solely dictated by combustion intensity attained in the bed and the space needed for the heat transfer surface. Factors such as (a) the amount of 'freeboard' needed above the bed to minimise elutriation, (b) size and location of steam headers and (c) arrangement of hot gas ducting can also have an important effect on the overall size of the plant.

Clearly the minimum size of containment will be obtained with the particle size range that gives the highest fluidising velocity; the conditions as regards heat transfer unfortunately are the converse of this.

Heat Transfer - The total heat transfer coefficient between a fluidised bed and an immersed surface is primarily a function of particle size, but it is also influenced by the temperature of the bed and of the immersed surface, and by the ability of bed material to circulate freely (i.e. on the closeness of tube packing). The total heat transfer coefficient, which comprises radiative and convective components, can be up to 10 times higher than in conventional gas-to-surface heat exchange systems, depending on particle size (Fig. 4). For example, in fluidised combustors burning 1/16"-0 coal, heat transfer coefficients to water cooled tubes of approx. 80 Btu/ft^2 h °F are obtained with tubes 2 in.apart, rising to 100 Btu/ft^2 h °F with tubes 6 in. apart. As Fig. 4 shows, higher heat transfer coefficients are obtained with higher tube temperatures.

A further factor that adds to the saving in heat transfer surface obtained with fluidised combustion is that the whole of the surface of the tubes immersed in the bed (and these absorb c. 70% of the heat of the fuel) is available for heat transfer, whereas only half of the surface of the tubing of a conventional boiler furnace is exposed to the combustion gases.

Operation under pressure does not affect the heat transfer to tubes immersed in the bed. Nevertheless, further savings in high-alloy tubing are achieved by operation under pressure, because the gas turbine absorbs most of the heat required to reduce the temperature of the gas leaving the bed to 700-750°F (375-400°C). The estimated overall savings in tube surface exposed to temperatures above 750°F (400°C), as compared with conventional p.f. firing for a 120 MW boiler, can be seen in Fig. 5. Although the pressure process requires a larger, more expensive economiser to recover heat below 750°F (400°C), this is a small consideration compared with the savings in cost for the higher alloy tubing achieved with fluidised combustion.

Although the heat exchange surface required in fluidised beds burning 1/16"-0 coal is about half that for a coal feed of 1/4"-0, the tube savings thus obtained have to be set against the higher containment costs for the five-fold increase in bed plan area that would be needed.

For boilers in large central power plant, with advanced steam conditions, particularly those operated under pressure, the choice may be for the finer coal size because of the major savings in high temperature alloys that would accrue. Industrial boilers operated at atmospheric pressure however, would need little or no high alloy tubing since advanced steam conditions would seldom be used; for these it is probable that the savings in space and containment costs would make coarse fuel the logical choice.

In either event, fluid bed combustion can result in big savings in tubing requirements and the compact nature of the tubing and of containment that can be achieved would facilitate a maximum amount of factory prefabrication, with consequent savings in erection costs.

<u>Bed Temperature</u> - At the temperatures currently favoured for fluidised combustion systems in the U.K., the vapour pressures of the alkali components of ash known to play a part in fouling and corrosion of heat transfer surfaces are several orders of magnitude lower than at the temperatures in conventional combustion systems, and as a result sodium concentrations in the gases from fluid bed combustors are about 1 ppm, equivalent to less than 1% of the sodium content of the fuel. This is in contrast to the 10% that has been recorded with p.f.-fired boilers, or 38% for a cyclone-fired boiler (Ounsted, 1958). The potassium contents of gases from fluidised bed combustors are typically less than those of sodium.

Combustion of residual oil in "chemically active beds" can result in a similar dramatic reduction in the concentration of fouling constituents in the combustion gases; deposition compounds of sodium, vanadium, and sulphur on superheaters has led to even greater restrictions on superheat temperatures for oil firing than for coal firing. Fluidised combustion may therefore be the key to the resumption of advances in superheat temperature for both coal and oil firing, and hence to realising the improvements in power generating efficiency predicted in the past (e.g. Downs, 1955).

For fluidised combustion under pressure, the low concentration of fouling constituents in the combustion gases gives hope that the gases can be expanded through gas turbines without loss of performance. In the past, successful operation of solid fuel fired gas turbines has been prevented by blade fouling or, in the extreme, by choking of the passageways through the blade system.

Erosion of coal-fired gas turbines has also been a problem in the past, and from this point of view low bed temperature is also an advantage; ash particles from a bed operating at 1470°F (800°C) show no signs of fusion and are similar in texture, appearance and composition to those produced by ashing in conventional laboratory ashing furnaces. In addition to being less abrasive than normal p.f. combustor ash, fluid-bed ash is coarser and is simpler to separate from the combustion gases; there is therefore good reason to believe that turbine blade erosion will not be a problem.

If, however, the low fouling and erosion propensities of the gases could be sustained to appreciably higher bed temperatures than 1470°F (800°C), additional advantages would accrue, e.g. (i) a further reduction in heat transfer surface, (ii) higher gas turbine efficiency, and (iii) simpler conditions for achieving a high combustion efficiency when burning the coarser grades of coal. Higher bed temperatures may ultimately be feasible for non-pressurised systems, but at the present state of the art it would not be prudent to predict this possibility for pressurised systems, particularly as there is evidence to show that particle temperatures can be appreciably higher than the mean bed temperature.

Bed temperature is also important from the point of view of emission/retention of the oxides of sulphur and nitrogen. At 1470°F (800°C) and thereabouts, the concentration of oxides of nitrogen may be less than 100 ppm.

Emission of sulphur as SO_2 can be reduced to less than 10% of the sulphur content of the coal (McLaren and William) by adding lime to the bed equivalent to rather less than twice the stoichiometric quantity; lime also reduces emission of nitrogen oxides.

There is greater tolerance of bed temperature from the point of view of sulphur retention (Fig. 6) however, than there is for volatilisation of alkalis.

3. SOME APPLICATIONS FOR FLUIDISED COMBUSTION UNDER PRESSURE

The power generation industry is a major consumer of coal and residual oil, and in spite of competition from nuclear power, it would be surprising if throughout the world fossil fuels failed to retain the greater part of the power generating market for at least another 20 years. Most of the work on fluidised combustion has consequently been carried out with this in mind, and almost all of the potential applications of pressurised fluid bed combustion discussed here are concerned with power generation.

3.1. Combined Power Generation Cycles

The thermal efficiency of power generation processes is mainly determined by the practical limitations on the temperature range over which the working fluid converts heat energy to power. The properties of ideal working fluids have been specified by Meyer and Fischer (1962) and by Reti (1965), who concluded that no known fluid possesses all the required properties. Steam, however, has more of the desired properties than any other single fluid.

The problems of increasing the temperature of high pressure steam have already been mentioned. There is little prospect of reducing low-pressure steam turbine exhaust temperatures below their present levels of around 80 - 90°F or 25 - 35°C without resorting to uneconomically large condensers and cooling water flows. The temperature range and efficiency of conventional cycles have therefore reached their limits, particularly for smaller plant.

Combined power generation cycles increase the working temperature range by using more than one fluid. The literature abounds with proposals for employing "topping" fluids to increase the maximum working temperature, and "bottoming" to reduce the minimum working temperature. Many of these cycles would need a clean fuel as heat source to avoid the unacceptable corrosion, erosion and fouling that would occur with conventional combustion of p.f. or residual oil. There are, however, a number of cycles which, if pressurised fluid bed combustion were successfully developed, would be suitable immediately or in the future for coal or residual oil firing.

Combined Gas-Steam Cycles - There are basically two types:

a. Exhaust-fired boilers, in which the combustion air to a more or less conventional boiler is replaced by the hot but oxygen rich exhaust gas from a gas turbine power plant.

b. Supercharged boilers, in which the boiler furnace operates at high pressure and the combustion gases are expanded through a turbine that provides power to drive the combustion air compressor and an additional alternator.

Both systems have their advocates, and their thermodynamic and economic principles have been discussed by Seippel and Bercuter (1961) and by Sheldon and McKone (1962). The supercharged combined cycle appeared the most likely to provide a worthwhile improvement in efficiency and capital cost in-so-far as combustion of coal was concerned, and experimental work and thermodynamic studies at BCURA have concentrated on this system.

An arrangement of a supercharged boiler cycle, based on a compression ratio of 8:1 and a gas turbine entry temperature of 1380°F (750°C) is shown in Fig. 7. Because the combustion air is heated by compression to over 500°F (approx. 250°C), alternative ways have to be found for recovering heat from the turbine exhaust. In the arrangement shown in Fig. 7, some of the feed heaters of the standard 120 MW steam cycle have been replaced by a low level economiser. Although the reduced extraction of steam for feed heating gives a poorer steam turbine heat rate, the power output is increased by about 5 MW. The calculated heat rate for such an arrangement (8580 Btu/kWh), thermal efficiency 39.8% based on the gross calorific value of the coal) would be about 450 Btu/kWh (about 2 percentage points in thermal efficiency) better than for a conventional steam plant.

Operation over a wide load range should be feasible, a feature of particular advantage for sites not served by power distribution networks.

This system, though not offering dramatic improvements in efficiency, would provide a good starting point for introducing the power industry to advanced cycles, and to fluidised combustion of coal and residual oil. A particular advantage of the combined cycle over pure gas turbine cycles, from the point of view of using these fuels, is that any fall off in turbine power from deposition or erosion has a much smaller effect on the output of the whole plant.

Gas - Potassium - Steam Cycle - In this cycle, potassium would be used for "topping" a conventional steam cycle by combining the functions of a potassium condenser with those of a steam boiler, superheater and reheater. A pressurised fluid bed combustor would provide a favourable heating system for generating the potassium vapour, since fireside corrosion would be less than with other combustion systems. Heat rates of 6250 Btu/kWh (thermal efficiency 54.6%) have been projected by Frass (1966) for a system with nuclear heating of potassium vapour to temperatures of 1540°F (838°C) at 29 psia. Taking into account the currently envisaged fluid bed temperatures; together with stack-losses, recalculation of the data gives a heat rate of about 7100 Btu/kWh (thermal efficiency 48%). Though this is a large improvement over conventional plant, it might not compensate for the higher capital costs likely to be incurred, and this challenging system must await the next generation of fluid bed combustors operating at higher temperatures.

Refrigerant Cycles - The use of refrigerants as "bottoming" fluids has been suggested by a number of authors for reducing the capital cost or improving the efficiency of steam cycles (e.g. Aronson, 1961) and gas turbine cycles (Hicks, 1965; Bindon and Carmichael, 1968). Based on an exhaustive analysis, Eaves and Hadrill (1968) concluded that, for U.K. conditions with conventional combustion systems, both capital and operating costs of combined cycles using refrigerants were unattractive. The situation might be changed with pressurised fluid bed combustion; by using lime to remove sulphur, low stack temperatures giving higher efficiencies would be possible without risk of low-temperature corrosion.

3.2. Pure Gas Turbine Cycles

Pressurised fluid bed combustion potentially converts "dirty" fuels to acceptably "clean" fuels. In addition to its application to combined cycle power plant, therefore, it has attractions for coal or residual oil firing of pure gas turbine cycles which at present need clean fuels, nuclear energy or large costly air heaters.

Simple Open Cycles - These are more susceptible to the effects of pressure loss than combined cycle plant. Attempts to develop directly-fired coal-burning gas turbines over the past two decades (Cox, 1951; Rozenberg, 1962; Wisdom, 1964; Nabors, Strimbeck, Cargill and Smith, 1965; Smith, Strimbeck, Coates and McGee, 1966) have been beset with difficulties of blade erosion and fouling by ash; these difficulties should be minimised by fluidised combustion. Simple open cycles are characterised by high air/fuel ratios, and to avoid excessive fluid bed cross sections, most of the excess air would need to be indirectly heated in tubes immersed in the bed (Fig. 8). Although fluidised combustion may successfully overcome the main problems that have prevented successful firing of turbines by coal and residual oil the number of applications where the additional cost of the combustion system would be justified may be limited.

Semi-Closed Cycles - These are characterised by having a high-pressure closed circuit, using relatively small, cheap turbomachinery, with a bleed of air or gas to an open cycle gas turbine operating at lower pressure. They have good start-up and part-load characteristics, and better heat rate (e.g. 10700 Btu/kWh or 32% efficiency) than simple open cycle plant, (Gasparovic, 1967) but their main potentiality is in the field of peak-load power generation. Pressurised fluid bed combustion could be used in semi-closed cycles in a number of ways which can be categorised as :

(a) Cycles with direct firing of the closed circuit (Fig. 9)

(b) Cycles with indirect heating of the closed circuit and direct heating of the open circuit (Fig. 10)

Closed Cycles - Several p.f.-fired closed air cycle plants have been constructed (Keller and Gaehler, 1961; Auer, 1961). In some of these, the heat rejected by the air before recompression was used for works or district heating, giving high heat utilisation efficiency. The air heaters were large and costly (Bammert and Nickel, 1966). Major savings should be possible with pressurised fluid bed combustion for heating the air and generating additional power (Fig. 11). Helium is a better closed cycle fluid than air because of its higher specific heat and ratio of specific heats at constant pressure and volume, its lower molecular weight, its higher thermal conductivity and its chemical inertness. A 25 MW nuclear heated helium closed cycle plant is being constructed at Geesthacht, W. Germany (Keller and Schmidt, 1967). Its heat rate of 9220 Btu/kWh (37% thermal efficiency) with a helium turbine entry temperature of 1350°F (730°C), could probably be closely matched if the nuclear heat source were replaced by a pressurised fluid bed combustor.

Air Storage - Fluidised combustion under pressure would be a suitable means for firing air storage power schemes (Stal-Laval) for peak load power production. The arrangement would be as shown in Fig. 12. Air, compressed to approx. 600 psia using off-peak electrical power, would be stored in a subterranean cavern under a constant head provided by a lake, reservoir or sea. Peak power would be generated by an indirectly heated gas turbine followed by a directly fired gas turbine system, with a heat consumption of 4350 Btu/kWh (thermal efficiency 79%)

3.3. Continuous Reforming of Hydrocarbons

Reforming of liquid and gaseous hydrocarbons is a major feature of processes for making ammonia, ethylene and town gas. 'Reforming' is probably a misnomer because, as is well known, the process taking place in the heated tubes of a continuous reforming unit is largely catalysed thermal decomposition of higher hydrocarbons to a mixture of lower hydrocarbons, carbon oxides and hydrogen. The relative proportions of the gases produced depend upon the temperatures and pressures that are used.

The equilibrium temperatures generally range from 1200°F (650°C) for manufacture of town gas to over 1800°F (1000°C) for production of ammonia synthesis gas and hydrogen. Tube wall temperatures required in reformer units, which could be heated by fluidised combustion furnaces, would therefore be in the range 1400°F (750°C) to 2000°F (1100°C).

The reforming equilibria are favoured by operating at high pressure; for example, the yield of methane can be increased. Metallurgical considerations are of major importance in design of the plant, since at the pressures (up to 600 psi) and temperatures used the tube metal creeps. By operating a fluidised combustion furnace under pressure, the pressure difference across the tube walls could be reduced or eliminated. Capital costs could thereby be reduced because thinner-walled tubing could be used, the materials being chosen mainly with regard to resistance to corrosion. Maintenance costs could also be lowered, since the reduction in creep would give a large (e.g. tenfold) increase in tube life.

4. DEVELOPMENT OF THE COMBUSTION SYSTEM

In a system that has so many potential advantages over other combustion systems for coal and residual oil there are inevitably some problems. We do not know of any that are likely to be unsolvable, but an appreciable amount of development work has been done, and more may be needed to ensure that the solutions do not detract significantly from the savings in cost that the benefits of the system should bring.

A major research effort is in progress at the Leatherhead research laboratories of BCURA, and the Cheltenham research laboratories of the NCB, aimed at proving the 'virtues' and overcoming the potential 'vices'. The pilot-scale equipment includes (a) a pressurised combustor with a bed area of 8 ft^2 capable of burning 500 lb/h of 1/16"-0 coal at 5 atm pressure, (Fig. 13); (b) a boiler with a bed area of 12 ft^2 capable of burning 1200 lb/h of ¼"-0 coal at atmospheric pressure; and (c) three other pilot-scale combustors capable of carrying out long term (e.g. 1000 h) test programmes.

4.1. The Programme

These research programmes include in their objectives:
(a) minimising the number of coal injection points needed to obtain good distribution without the need for excessively deep beds. Deep beds entail high pressure loss, and the larger the number of coal injection points the more expensive is the distribution system;
(b) evolving the optimum means for recycling incompletely burned particles so as to obtain a high combustion efficiency. The design of the space above the bed is also important from this point of view;
(c) exploring means for improving dust collector performance. Two stages of dust collectors are used on the pilot-scale pressurised combustor. The cleaned gases contain only a small proportion of particles larger than 10 μm, and the concentration of dust passing over the cascade downstream of the dust collectors is much lower than for previous solid fuel fired gas turbines. Although the pressure loss over the cleaning system is acceptable, lower pressure loss is desirable;
(d) establishing the best procedures for part load operation. Operating over a wide load range in a system in which the main factor that controls heat absorption, namely bed temperature, can only be varied over a small range presents a certain amount of difficulty. Two main methods have been followed (i) to stop fluidising parts of the bed (i.e. to compartment the bed) and (ii) to expose part of the tubing to gas from the bed by altering bed level (heat transfer coefficients to tubes in the bed are several times those to tubes above the bed).

Operation over a wide load range in the combined cycle will require some care in matching combustion and heat transfer characteristics of the fluid bed to the pressure, temperature and mass flow characteristics of the gas turbine. Fig. 14 shows the part-load characteristics of a typical industrial gas generator and Fig. 15 shows the calculated effects on air-fuel ratios (overall and in the bed) of following one method of load control. Achieving this should not involve an excessively expensive control system.

An expanding part of the programme involves a comprehensive series of investigations to optimise operating conditions with additions of lime (as limestone or dolomite) to the bed to reduce atmospheric pollution by sulphur and nitrogen oxides. An agreement between the NCB and the National Air Pollution Control Association provides for a full exchange of information on these and other aspects of the research programme.

4.2 The State of the Art

The pilot-scale pressure combustor has been operated successfully for several hundred hours. Combustion efficiencies of 99% can be achieved, and few operating difficulties have been experienced with the fluid bed.

Heat transfer rates to tubes in the bed are similar to those expected from tests at atmospheric pressure. The combustion gases at approx. 1400°F (750°C) have been passed over two designs of static turbine blade cascade; that currently in use is a segment from the first stage stator row of a marine version of an aero engine. The approach velocity is about 400 ft/s and the leaving velocity about 1800 ft/s. No erosion of the blades, or of a target tube downstream of them has occurred. Very little ash has deposited on the surfaces, and deposits that have formed can be easily removed by 'on line' cleaning methods commonly used for gas turbine compressors.

There is cause for cautious optimism as to the technical success of the process. The next stage of proving this involves long term running of a gas turbine on gases from a fluidised combustor. This will be an expensive project and one of many pre-requisites is for design studies to show that the process can be justified economically.

The size and the characteristics of the gas turbines likely to be available in the next 5 - 10 years could be a dominant factor, because the cost of developing turbines specially for the process would be prohibitive. High pressure ratios (e.g. 15 : 1), good part load characteristics, and air mass flows of 300 lb/s upwards are amongst the features desired.

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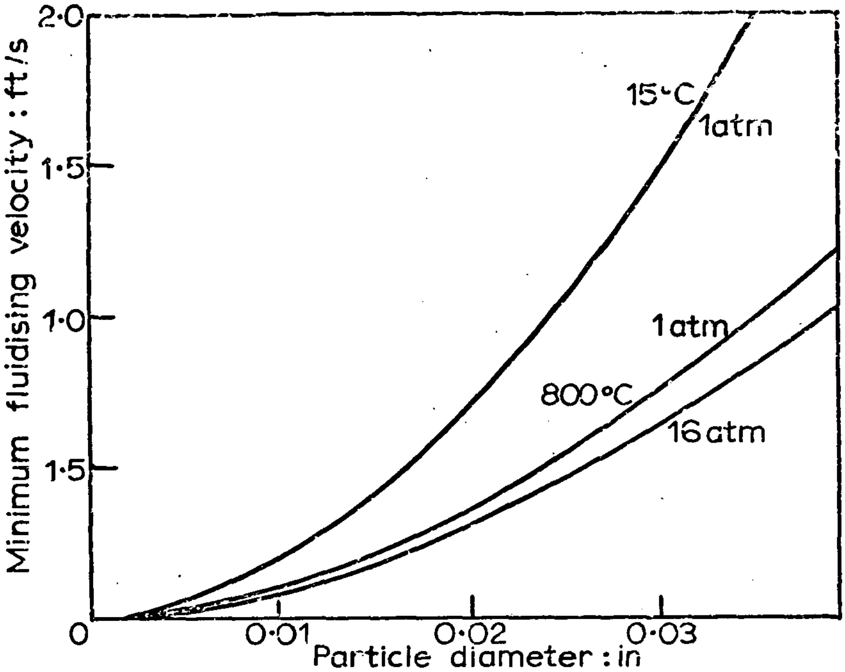

Fig. 1. Minimum fluidising velocity : effect of pressure.

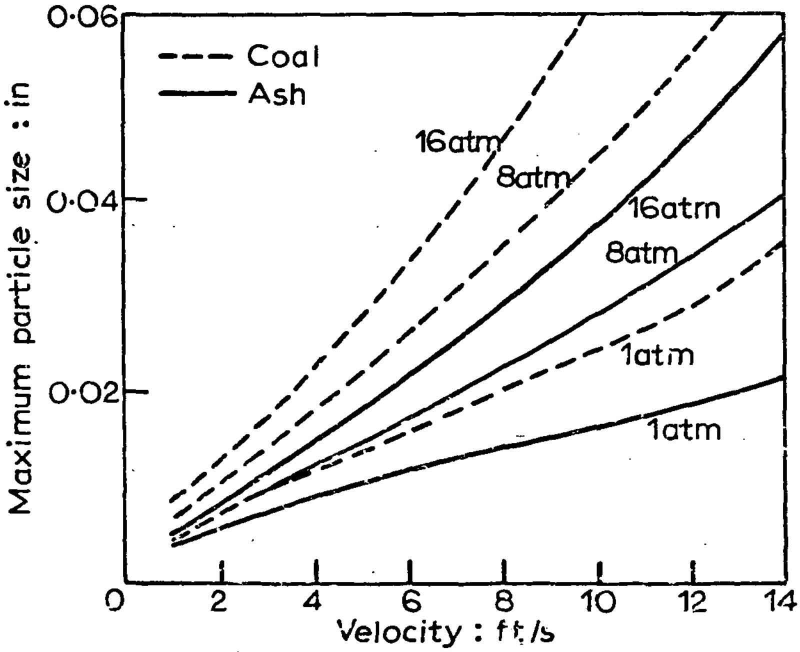

Fig. 2. Size of particle elutriated : effect of pressure

120/1 mh

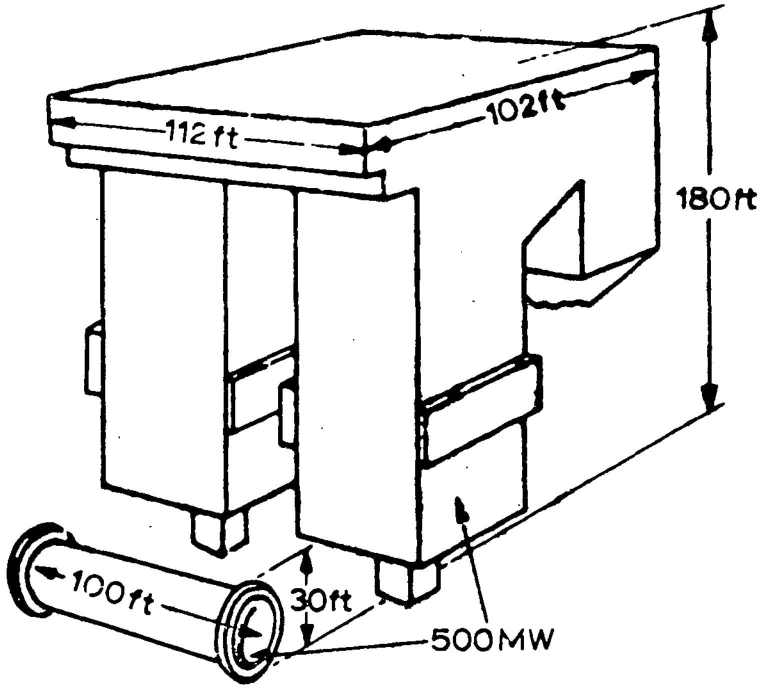

Fig. 3. Comparison of boiler size : pressurised fluid bed v. conventlal p.f.

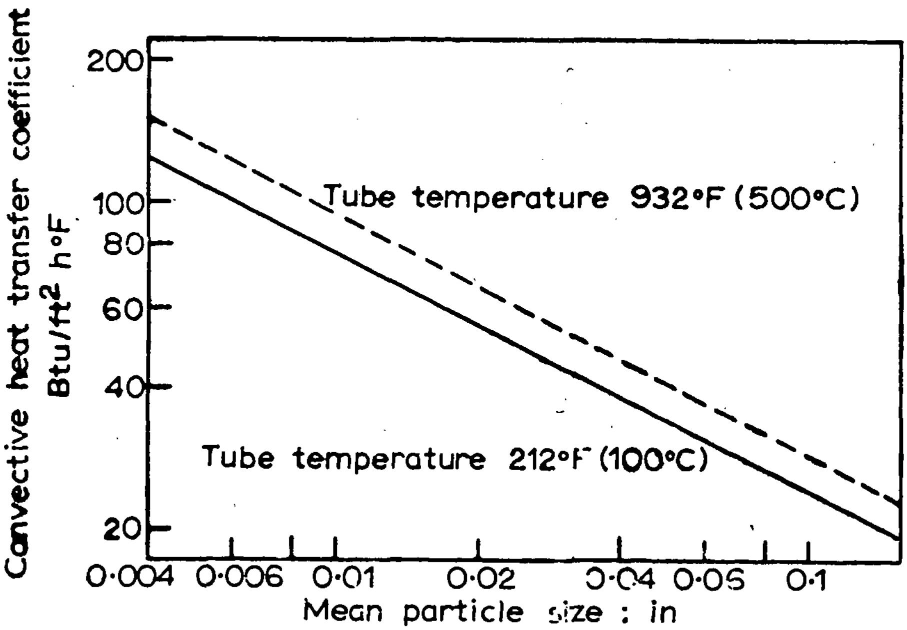

Fig. 4. Fluid bed heat transfer coefficients

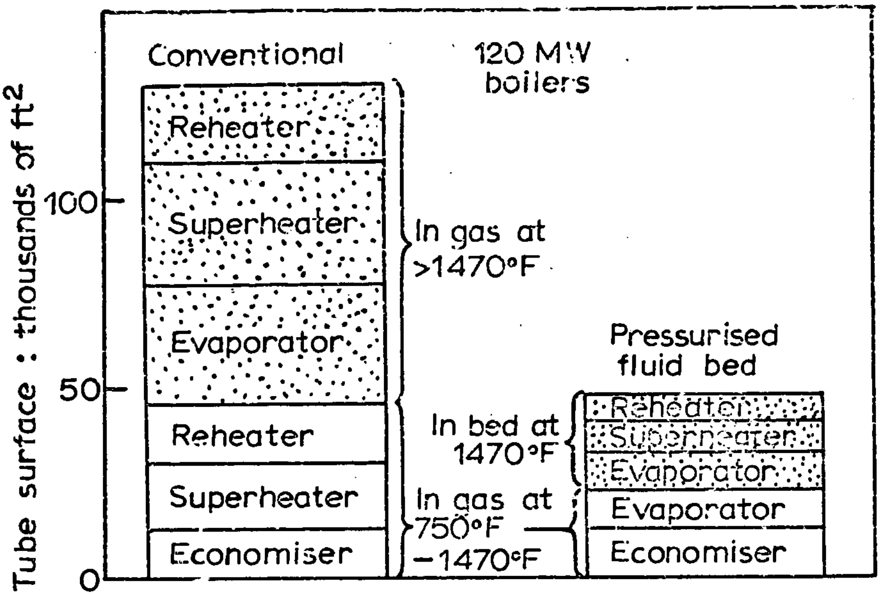

Fig. 5. Comparison of tube surface areas

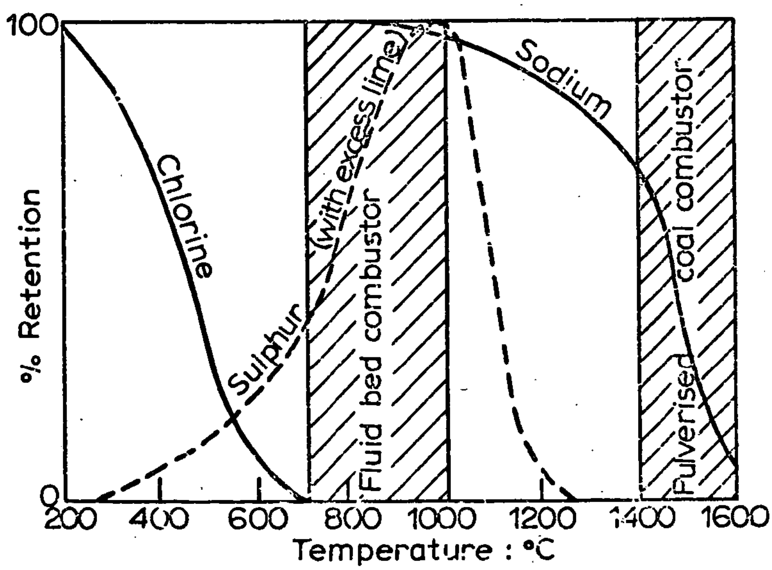

Fig. 6. Retention of ash constituents

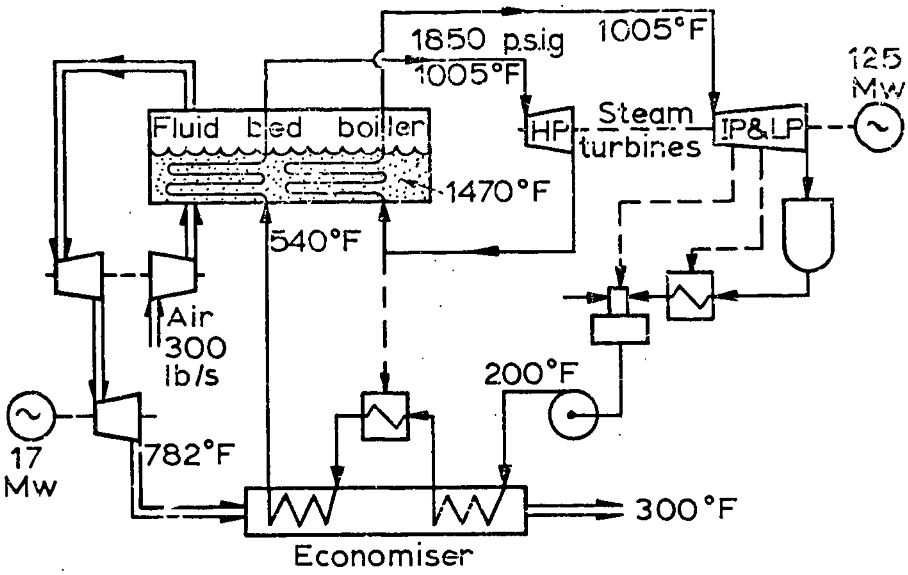

Fig.7. 140 MW Gas-steam turbine system

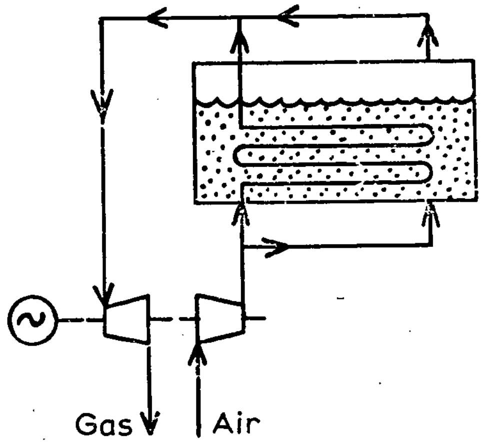

Fig. 8. Open cycle plant

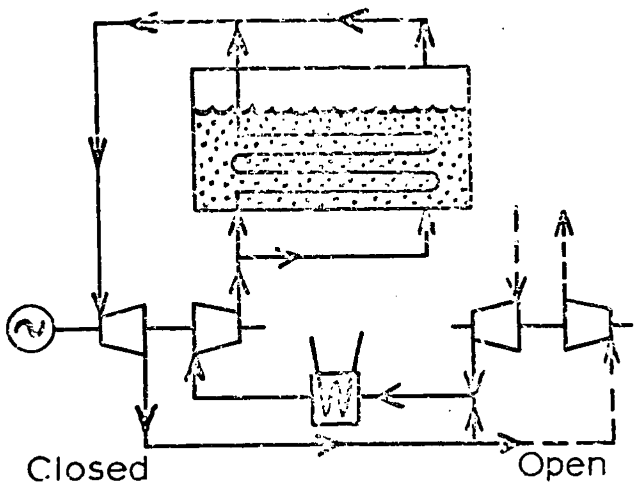

Fig. 9. Semi-closed cycle : combustion in the closed cycle

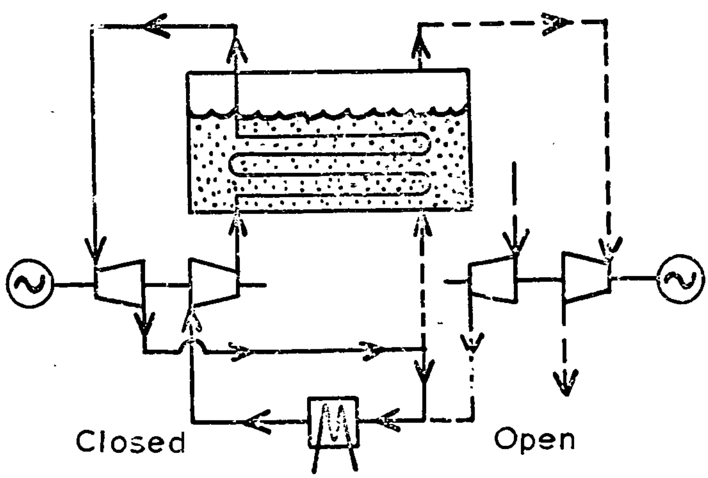

Fig. 10. Semi-closed cycle : combustion in the open cycle

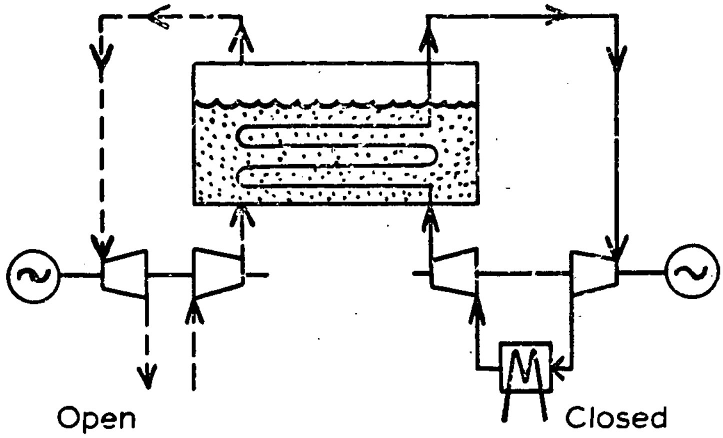

Fig.11. Closed cycle plant

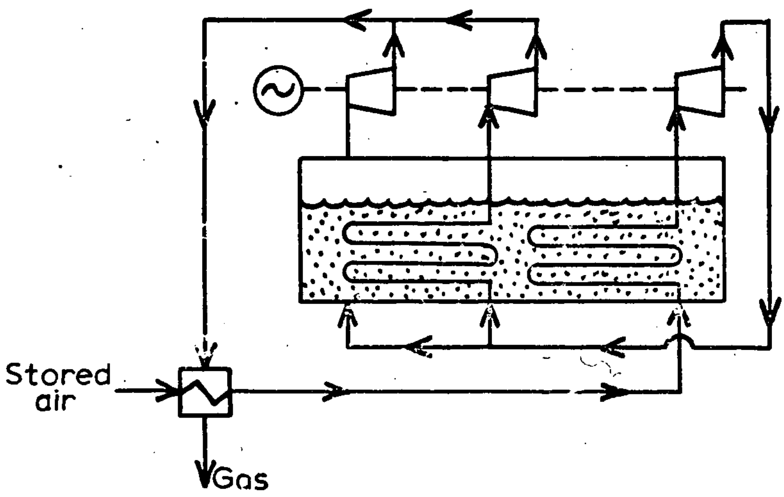

Fig. 12. Air storage scheme

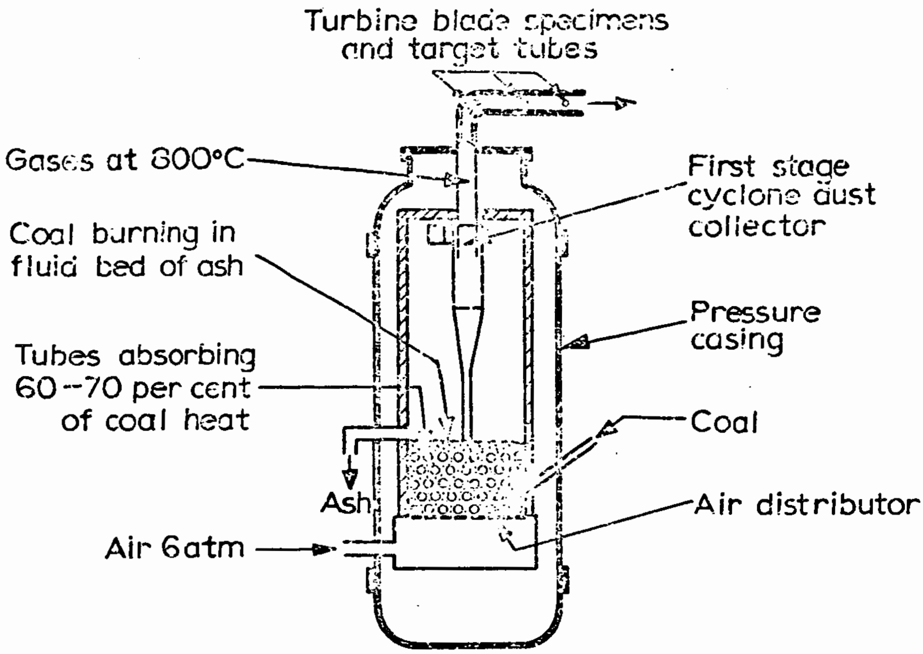

Fig. 13. Experimental pressurised fluid bed combustor

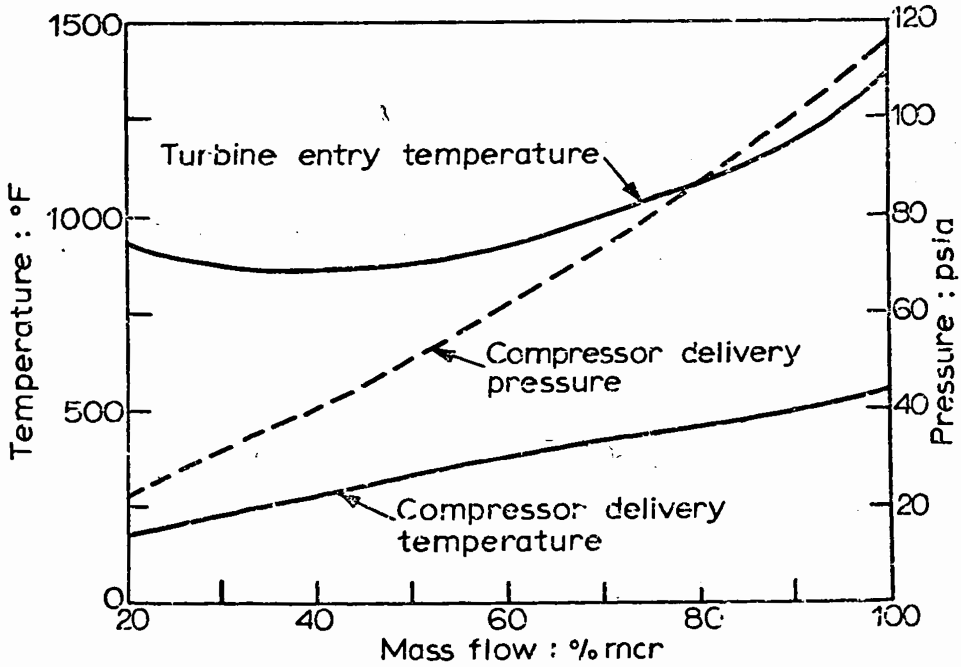

Fig. 14. Part load characteristics: industrial gas generator

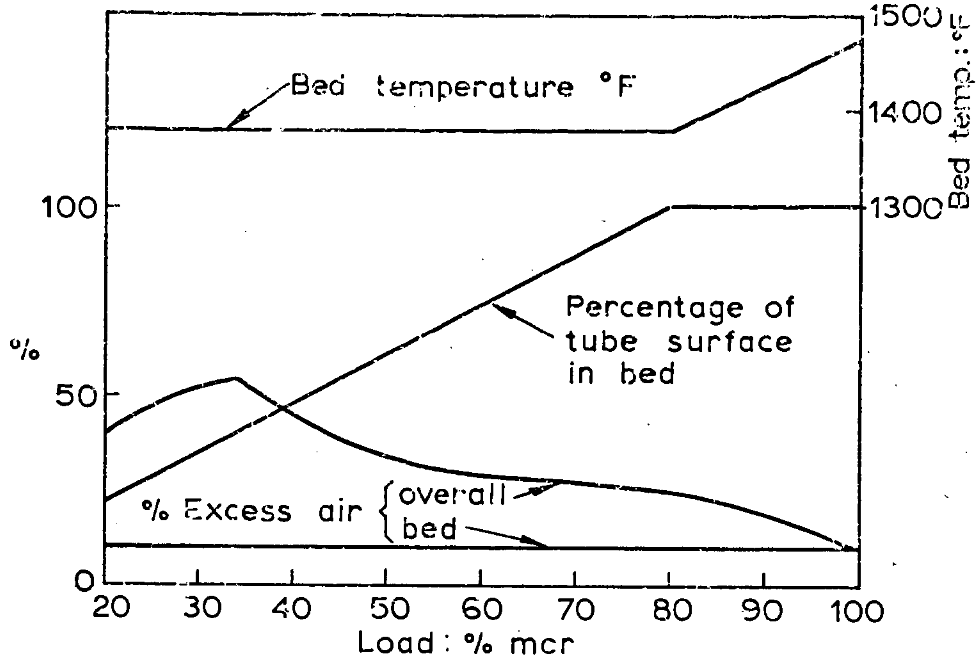

Fig.15. Calculated part-load characteristics : combustor with variable bed level

PROCESSING COAL FOR POWER GENERATION BY MHD

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Office of Coal Research

Processes for gasifying and/or pyrolyzing coal to produce superior fuels for power generation by MHD are presented. Also several energy complexes where MHD power generation is mated with the conversion of coal to synthetic crude oil and high Btu gas are described.

Advanced-Cycle Power Systems Utilizing Desulfurized Fuels

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INTRODUCTION

The electric utility industry in the United States currently contributes approximately 50% of the nearly 30 million tons per year of sulfur oxides emitted into the atmosphere (Ref. 1). Since the total installed capacity of electric utilities is projected to double each decade, (Ref. 2) the amount of sulfur oxides emitted into the atmosphere in future years could exceed projected standards in some sections of the country unless suitable methods are developed to control sulfur oxide emissions. Many processes, ranging from cleanup of the stack gas to cleanup of fuel before combustion, have been proposed for controlling sulfur oxide emissions from power plants. Although many stack gas cleaning methods are technically feasible, most of them are expensive and have not demonstrated reliable operation in commercial service. The alternative approach, involving removal of sulfur from fuel before combustion, looks most promising as a long-range solution for controlling sulfur oxide pollution from fossil-fueled central power stations (Ref. 3).

The removal of sulfur from fossil fuels before combustion can be a difficult task, and the resulting fuel delivered to the power system is certain to cost more than the raw fuel which serves as feedstock. In order to evaluate the technical and economic feasibility of fuel desulfurization processes as an alternative to stack gas desulfurization, it is necessary to reappraise the traditional methods of electric power generation and to evaluate advanced power systems which may be capable of operating at higher efficiencies than conventional steam systems. The Research Laboratories of United Aircraft Corporation, under contract to the National Air Pollution Control Administration*, are currently investigating the technical and economic feasibility of desulfurizing coal and residual oil and the utilization of these desulfurized fuels in advanced-cycle power systems. This paper describes the results of cycle analysis of various advanced power systems including preliminary estimates for the cost of generating power with these systems and indicates the benefits that may arise through the use of desulfurized fuels.

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* Contract No. CPA 22-69-114

FUEL PROCESSING

While this paper deals exclusively with the preliminary results of a study on advanced-cycle power systems, it is appropriate to discuss briefly at this time the type of fuel used in this investigation. In order to meet sulfur oxide emission control standards, the sulfur contained in coal or residual oil must be significantly reduced or removed altogether. One method of obtaining a low-sulfur fuel is through the partial oxidation of coal with steam and air at high temperature and pressure. The product gas from this process, with sulfur now in the form of H_2S and COS is then sent to a desulfurization unit which removes the sulfur compounds. Since the gasification processes and desulfurization techniques of interest would typically operate at elevated pressures, a high-pressure fuel gas having a heating value of 150 to 200 Btu/ft^3 at standard conditions would result. Such a fuel gas has a number of advantages in combination with the power systems to be discussed, and therefore this type of fuel was assumed in all the systems investigated.

Preliminary estimates of the cost of potentially attractive gasification processes have indicated that a clean, desulfurized fuel could be delivered to a mine-mouth powerplant for a cost which is 25% to 50% greater than the cost of the raw fuel.

ADVANCED-CYCLE POWER SYSTEMS

The specific types of advanced-cycle power systems investigated are shown in Table I, and can be grouped into two generic classifications: external-combustion power systems in which boilers or heaters are used to heat the working fluid, and internal-combustion power systems in which the products of combustion constitute the working fluid. The external-combustion systems investigated include those using the conventional steam cycle, binary cycles utilizing steam and other working fluids, and cycles such as the closed-cycle gas turbine in which the working fluid is heated in a gas heater. The internal-combustion systems studied were essentially based on variations of the gas turbine cycle and included consideration of power systems using the combined gas and steam (COGAS) turbine cycles. The investigations were based on present-day power system technology, although possibly not yet reduced to commercial practice, as well as technology judged to become commercially available in the 1980 and 1990 decades.

Steam Systems

Conventional steam power systems were included in the investigation to provide a basis of comparison for all other power systems. Performance estimates predicted for projected future conventional steam power systems are given in Table II. It is apparent from this table that the overall station efficiencies for these systems are not projected to increase substantially in the time span considered, because technology available for use in steam power stations has reached a plateau. Relatively minor increases in station efficiency will be possible due to slight increases in the internal efficiencies of turbogenerators and boilers, but substantial increases in efficiency due to improved cycle conditions are not foreseen, since any increase in cycle temperature and pressure would result in a very marked increase in system capital costs. Thus, no significant increase in conventional steam power station thermal efficiency is foreseen to offset the increased costs projected for future desulfurized fuels.

Binary Cycles

The basic steam power station may be modified by the addition of a binary cycle to either increase its efficiency or decrease its capital cost, both with the goal of maintaining or reducing the cost of power. One method which has been suggested (see Ref. 4, for example) for reducing the capital cost of power stations is to stop the steam expansion at approximately 35 psia, eliminate the relatively expensive low-pressure sections of the steam turbine, and incorporate an ammonia or fluorocarbon bottoming cycle which would operate at relatively high pressure. Supposedly, the capital cost of the bottoming system would be less than for the steam equipment it replaces. A temperature-entropy diagram for a 3500 psig/1000 F/1000 F steam cycle with an ammonia bottoming cycle is depicted in Fig. 1. The efficiency of a power station incorporating this type of a bottoming cycle would be substantially less than that of a 3500 psig/1000 F/1000 F steam station, i.e., approximately 35.7% compared to 38.6%, because of the irreversible heat transfer between steam and ammonia, and because it is estimated that the ammonia turbine would be capable of attaining a slightly lower isentropic efficiency than the section of the steam turbine which it would replace. Analysis has shown that the increased fuel consumption due to reduced cycle efficiency would more than offset any system capital cost reductions that might be anticipated with a bottoming cycle. In fact, when account is taken of various capital cost penalties associated with the use of fluids other than steam, such as the need to employ welded construction to minimize leakage, the total capital cost of a bottoming cycle would not be significantly different than the cost of the conventional steam equipment it would replace.

A method which has been suggested for increasing the performance of the power stations is to use a high-temperature topping cycle which would reject its heat in a steam boiler. Previous studies (Refs. 5 and 6) of this type of cycle have indicated that potassium would be the best topping-cycle working fluid. In order to eliminate the upper limit of system performance, the potassium topping cycle shown in Fig. 2 was analyzed. The potassium would be pumped to its highest pressure of only 152 psia, heated to 2000 F by combustion gases, expanded through a turbine equipped with moisture separators to keep the moisture content of the potassium from exceeding 12%, and finally exhausted into a steam heater wherein the potassium would be condensed. If such a potassium cycle were used in conjunction with a 3500 psig/1000 F/1000 F steam cycle, a binary cycle efficiency of 58.8% would result. The overall station efficiency, taking into account such factors as boiler losses, generator efficiencies, and pump losses, would be 50.6%. This value is more than 10 points higher than the efficiency of the best present-day steam plant. As in the case of the bottoming cycles, the cost of the system equipment would be penalized in comparison with conventional steam systems because of the necessity to eliminate leakage of the working fluid and also to provide safety equipment to minimize the potential effects of a potassium-water reaction. Also, the heat exchangers require very costly materials for construction. Thus, the estimated capital cost for this system of over $200/kw is significantly higher than that for conventional plants, and it is estimated that the total cost of producing electricity would not be reduced relative to the cost with conventional steam stations. Furthermore, the problems associated with the successful development of potassium turbines of several hundred megawatts capacity are very complex.

Closed-Cycle Systems

A second group of external-combustion cycles is formed by what could be called closed-cycle gas turbine systems. The systems investigated included a true Brayton-cycle system utilizing helium as the working fluid, a supercritical Rankine-cycle system utilizing CO_2 as the working fluid, and a combination Rankine and Brayton-cycle system utilizing SO_2 as the working fluid.

Several cycle configurations involving the use of intercooling, regeneration, and reheating were studied. Considerations of advanced materials suitable for use in the fluid heaters indicated that tube wall temperatures would have to be restricted to 1800 F or below if an acceptable equipment lifetime of at least 100,000 hr were to be obtained. Thus, it was decided that the maximum working fluid temperature would be limited to 1600 F, and cycle evaluations and equipment sizing were performed for this value. A second temperature level of 1200 F at which advanced, but currently available, materials could be used was also selected for evaluation. By investigating a number of configurations for these two temperature levels, the tradeoff between cycle efficiency and equipment capital cost could be estimated.

Helium closed-cycle gas turbine systems have been the subject of widespread interest (e.g., Refs. 7, 8, and 9) because of the potentially high cycle thermal efficiencies such as those shown in Fig. 3. The efficiencies shown are for a cycle having one intercooler and a 90% effective regenerator. For the 1600-F temperature limit, the efficiency would be approximately 47%, and at the 1200-F level, approximately 44%. This efficiency can be increased somewhat by going to a different configuration, and the system selected for further evaluation at 1600 F would use two intercoolers and a 90% effective regenerator to give an estimated 48.5% cycle efficiency. The resulting power station would have a net efficiency of 41% with an estimated installed cost of $170/kw.

The use of CO_2 as a working fluid has been investigated a number of times, Refs. 9 and 10 being prime examples. Because of its low critical temperature, 88.7 F, CO_2 cannot be used in a Rankine cycle since the minimum cycle temperature allowed by the available cooling water is approximately 100 F. A typical cycle using CO_2 is shown in Fig. 4 in which it is seen that the flow would be split into two streams, one being compressed in a gas compressor and the other being cooled to supercritical liquid and then pumped up to maximum cycle pressure. A configuration such as this reduces the total compressor work required and would also allow the use of extensive regneration. The configuration of Fig. 4 would result in an overall station efficiency of 39% at an estimated capital cost of about $200/kw.

The final working fluid considered for the external-combustion cycles was SO_2. While this fluid is toxic, it does exhibit other properties which make it an interesting fluid for power systems (Ref. 11). The critical temperature of SO_2 is 315 F; thus, a condensing cycle can be considered. The 1600-F cycle selected for evaluation is shown in Fig. 5. Like the CO_2 cycle, the flow would be split into two streams: one compressed as a gas, the second condensed to the liquid state and pumped to maximum cycle pressure. By utilizing a reheat cycle and a 92.5% effective regenerator, this cycle would exhibit an efficiency of nearly 59%. The overall station efficiency would be about 51% with an installed cost estimated to be \$227/kw, an appreciable portion of which can be attributed to the very large regenerator.

Thus, a number of variations of external-combustion cycles have been investigated with the intent of increasing efficiencies or decreasing capital cost in order to offset the potential increase in fuel cost which would result from using desulfurized fuel. The potential power costs for systems using these cycles are summarized in Fig. 6 in which the estimated generating cost in mills per kilowatt hour for each system is compared with that of a conventional steam system. The generating costs are given for two fuel costs, 30¢/million Btu, which is a typical price of present-day untreated residual fuel oil, and 50¢/million Btu, a price projected for typical future desulfurized fuels. The results presented in Fig. 6 show that none of the cycles discussed thus far demonstrate cost advantage over the conventional steam system.

Gas Turbine Systems

The second generic group of power systems considered for use in central stations consists of internal combustion systems in which the products of combustion constitute the working fluid. Contrary to the case for conventional steam systems in which no significant improvement in performance is foreseen during the time period of interest, industrial gas turbine technology is projected to continually improve during the next several decades, primarily because of fallout from advanced aircraft development programs (Refs. 12, 13, and 14). The use of aircraft gas turbine technology in large industrial-type gas turbines could then give rise to performance benefits that would allow these engines to become competitive with steam power systems.

Figure 7 lists some aspects of gas turbine technology for the three time periods under consideration. The projections shown in Fig. 7 indicate that both aerodynamic performance (i.e., compressor and turbine efficiencies) and turbine inlet temperatures increase with time. The projected improvements in turbine inlet temperature are due to two considerations: increases in materials technology, which allow blade materials to withstand higher operating temperatures, and improvements in blade cooling techniques. Historically, turbine inlet temperatures have advanced approximately 20 F per year because of materials improvements. This improvement is shown in Fig. 8 along with the improvement made possible by the use of several cooling techniques. Data points in Fig. 8 indicate actual or projected engines utilizing both improved materials and improved cooling techniques.

A major improvement in gas turbine performance could be realized if the compressor bleed air normally used to cool the turbine blades is precooled in an external heat exchanger to temperatures of about 125 F before being utilized in the turbine for cooling purposes. The performance improvements that would result from the use of precooled compressor bleed air are: (1) for the same amount of bleed air extraction, a higher turbine inlet temperature could be realized, or (2) a smaller extraction flow would be required to maintain a given turbine inlet temperature. The performance gains which might then be realized by using precooled bleed air are shown in Fig. 9, in which projected performance for three generations of engines is shown. Another benefit which might arise from the use of precooled compressor bleed air is that less costly impingement cooling might be used instead of transpiration cooling in very high-temperature engines.

The performance shown in Fig. 9 was based on the use of methane as fuel. The use of a fuel resulting from gasification of coal would actually improve the performance over that of an engine burning methane. This improvement is shown in Fig. 10 for an advanced-design engine. The improved performance results because the fuel gas supplied from a high-pressure (above 15 atm) coal gasification facility typically would have a low heating value (150 to 200 Btu/ft^3) and, thus, displace air which would normally be compressed in the compressor section of the gas turbine. The gas turbine would then operate at higher efficiency because there would be less compression work for the same net power output. The incentive to produce a clean, gasified fuel suitable for gas turbines is quite high since the use of such a fuel would allow the operation of gas turbine central stations which should be less costly than comparable steam stations and should operate at efficiencies equal to or better than those envisioned for steam power systems.

COGAS Systems

A second and potentially more promising system utilizing gas turbines is the combined gas and steam (COGAS) power system. A simplified schematic diagram for an exhaust-fired type of COGAS system is shown in Fig. 11. Fuel from the gasification process would be fed into the burner of a high-temperature gas turbine. After combustion and expansion through the gas turbine, the hot combustion products would be exhausted into a heat recovery boiler to raise steam for expansion through a steam turbine. Supplementary firing in the boiler would be optional. The application of COGAS power systems to large-capacity, base-load electric power generation (Refs. 15 and 16) has been limited primarily to the US Southwest where large quantities of low-cost natural gas are available. Even in this area, the small improvements in performance and cost offered by present-day COGAS systems relative to those of the conventional steam stations have not been sufficiently high to induce utilities to convert from conventional steam to COGAS systems. In those few large COGAS systems that have been put in operation, supplementary firing is employed in the boiler and the gas turbines produce less than 20% of the total station output.

Early results of this investigation indicated that station efficiency could be increased significantly if the amount of gas turbine participation were increased by reducing the amount of supplementary firing in the boiler. Further increases in COGAS performance would be possible by increasing the gas turbine inlet temperature. These trends are shown in Fig. 12 for methane-fueled COGAS systems incorporating a low-performance steam cycle (thermal efficiency of 34%) and gas turbine inlet temperatures of 2000, 2400, and 2800 F. The data in Fig. 12 clearly indicate that the best COGAS performance would be obtained if the steam boiler were of the simple waste-heat recovery type with no supplementary combustion. Detailed performance and economic analyses of various steam cycles for use in COGAS systems were carried out, resulting in selection of a 2400 psig/1000 F/1000 F steam cycle without feedwater heating. Performance estimates of COGAS systems using this steam cycle are shown in Fig. 13 for turbine inlet temperatures of 2000 F, 2400 F, and 2800 F, and for a range of applicable pressure ratios. These estimates are based on the use of both methane and a 162-Btu/ft^3 gas supplied at burner pressure. As with the simple gas turbine system, both cycle efficiency and power output per unit air-flow would be higher for the system burning low-Btu gas. The projected net station efficiencies of 50%, in systems using current technology, to 56% or 57% in later generations would be significant improvements over the efficiencies that might be realized by any other system except the very exotic and very expensive liquid-metal topping cycles. The COGAS system, however, would use machinery which is evolutionary in nature, i.e., machinery which is based upon actual power systems now being manufacgured.

By utilizing advanced cooling techniques such as precooled bleed air, the maximum turbine inlet temperature for the three time periods of interest are projected to increase to 2200, 2800, and 3100 F, respectively, resulting in efficiencies several points higher than those depicted in Fig. 13. The projected efficiencies of COGAS systems using precooled bleed air and burning low-Btu gas supplied at burner pressure and 150 F are shown in Fig. 14 to approach 58%, a value which is nearly 50% greater than now realized in conventional steam stations. This performance may be improved even more if the fuel gas were to be supplied to the system at a temperature higher than 150 F, as shown in Fig. 15 for a third-generation or 3100-F turbine inlet temperature system. At fuel gas temperatures of 600 F and above, the performance of the system could be boosted to values of 60% and over, a goal which should supply a tremendous incentive to fuel cleanup processes.

Preliminary estimates for the overall cost of electricity generated by a conventional steam system, a straight gas turbine system, and a GOGAS system are shown in Fig. 16. The projections presented in Fig. 16 demonstrate that the use of advanced technology in gas turbines could result in power systems which may produce electricity at costs equal to or even less than now realized from conventional steam systems, and still reduce the mission of sulfur oxides into the atmosphere. A second benefit arising from these systems is a reduction in thermal pollution of cooling water. The straight gas turbine rejects heat directly to the atmosphere; thus, there is no thermal pollution of cooling water. A reduction of thermal pollution by about 50% (compared to conventional steam stations) is possible with COGAS systems because of the increase in cycle efficiency, and because of the higher sensible heat content of the stack gases.

CONCLUSIONS

In conclusion, it can be said that the use of aircraft technology in industrial gas turbines may result in power systems which could produce electric power at reasonable cost using fuels which are appreciably more expensive than those used today, but which do not contain sulfur. The premise that advanced-cycle power systems could maintain the cost of producing electricity at levels now obtained with conventional systems has been shown to have great promise in future power systems. Additional benefits will occur through the use of advance-cycle power systems in areas of thermal pollution and in capital costs.

FUTURE WORK

Having determined the most promising generic classification of power systems, there remains a good deal of work to be done. Before detailed design work which would lead to actual engine development can be undertaken, further studies must be made with the objective of determining the best cycle configuration and operating conditions. Of immediate interest is the problem of combustion of the low-Btu fuel gas in the vitiated conditions occurring in a reheat combustor. The use of reheat in a COGAS cycle may allow significant gains in performance, but current limitations of funding do not allow a thorough study of this cycle. Operation of advanced-cycle power systems during transient periods will also require more detailed analyses, particularly in combination with the fuel gasification systems. Evaluation of the effect of these areas on system costs must also be made. Comparable work in the fuel clean up processes must also be performed, particularly in the area of high-temperature, high-pressure desulfurization techniques.

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TABLE I

ADVANCED-CYCLE POWER SYSTEMS INVESTIGATED

External-Combustion Systems

- Conventional Steam
- Binary Cycle Bottoming
- Binary Cycle Topping
- Closed-Cycle Gas Turbine

Internal-Combustion Systems

- Open-Cycle Gas Turbine
- Combined Gas and Steam (COGAS) Turbine

TABLE II

SUMMARY OF PERFORMANCE, AND CAPITAL AND GENERATING COSTS FOR CONVENTIONAL STEAM-ELECTRIC STATIONS

Fuel	No. Units and Size - Mw	Steam Conditions	Net Station Efficiency[1]	Capital Cost $/kw[2]
		PRESENT-DAY		
Untreated Coal	2-500	2400 psia/1000 F/1000 F	36.6%	$176.9
Gasified Coal	2-500	2400 psia/1000 F/1000 F	37.5	148.0
Untreated Oil	2-500	2400 psia/1000 F/1000 F	37.0	163.5
Desulfurized Oil	2-500	2400 psia/1000 F/1000 F	37.5	163.5
		SECOND-GENERATION		
Untreated Coal	1-1000	3500 psia/1000 F/1000 F	38.6	$162.7
Gasified Coal	1-1000	3500 psia/1000 F/1000 F	39.5	136.9
Untreated Oil	1-1000	3500 psia/1000 F/1000 F	39.0	150.1
Desulfurized Oil[3]	1-1000	3500 psia/1000 F/1000 F	39.3	150.1
		THIRD-GENERATION		
Untreated Coal	1-1000	3500 psia/1000 F/1000 F	39.6	$162.7
Gasified Coal	1-1000	3500 psia/1000 F/1000 F	40.5	136.9
Untreated Oil	1-1000	3500 psia/1000 F/1000 F	39.5	150.1
Desulfurized Oil	1-1000	3500 psia/1000 F/1000 F	40.5	136.9

(1) At a 70% annual load factor
(2) This is the installed cost for a station needing no unusual site preparation or station aesthetics.
(3) Residual oil may be gasified to meet second-generation requirements; thus the performance, capital costs, and generating costs for gasified oil-fired stations will be similar to those for the gasified coal station.

FIG. 1

AMMONIA BOTTOMING CYCLE

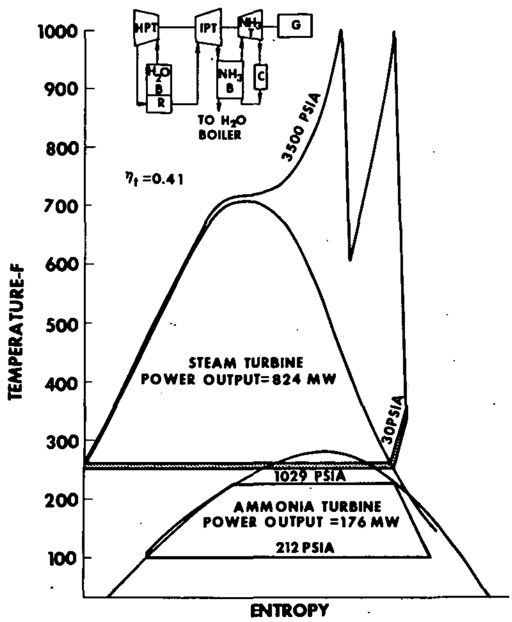


FIG. 2

HIGH-PERFORMANCE POTASSIUM TOPPING CYCLE

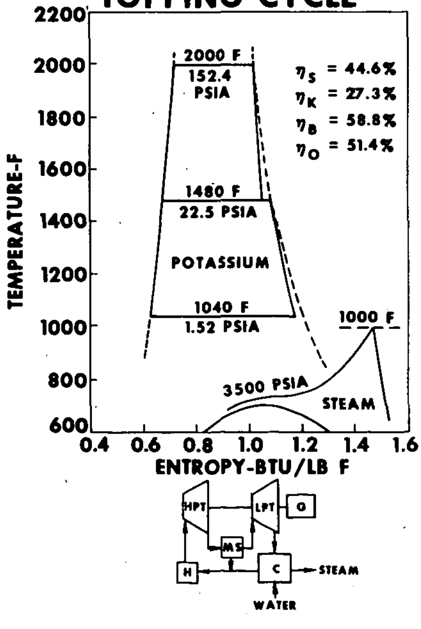


FIG. 3

HELIUM CLOSED-CYCLE TURBINE PERFORMANCE

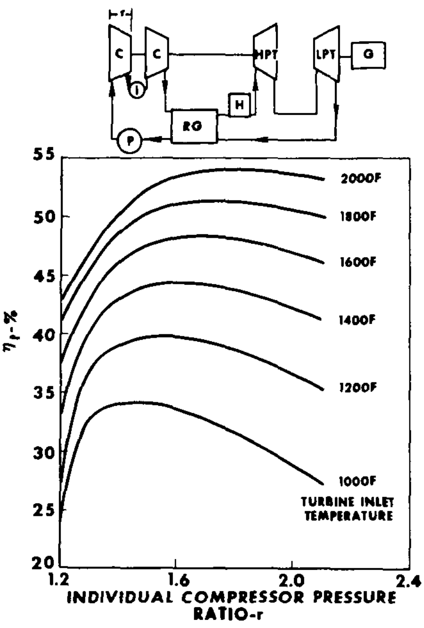


FIG. 4

CARBON DIOXIDE POWER CYCLE

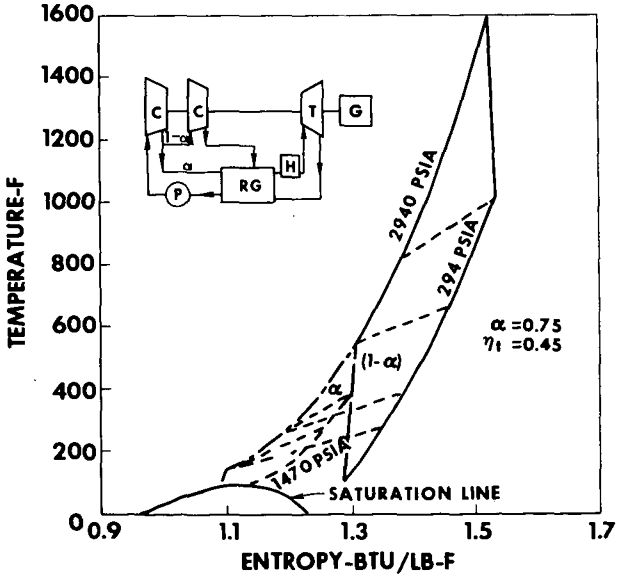


FIG. 5

HIGH PERFORMANCE SO_2 POWER CYCLE

REGENERATION EFFECTIVENESS = 92%
CYCLE THERMAL EFFICIENCY = 58%

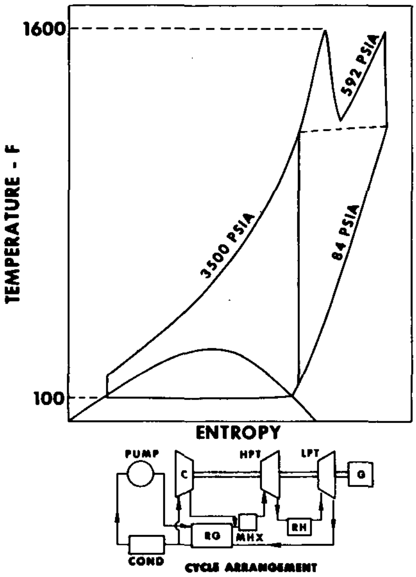


FIG. 6

COMPARATIVE ENERGY COSTS

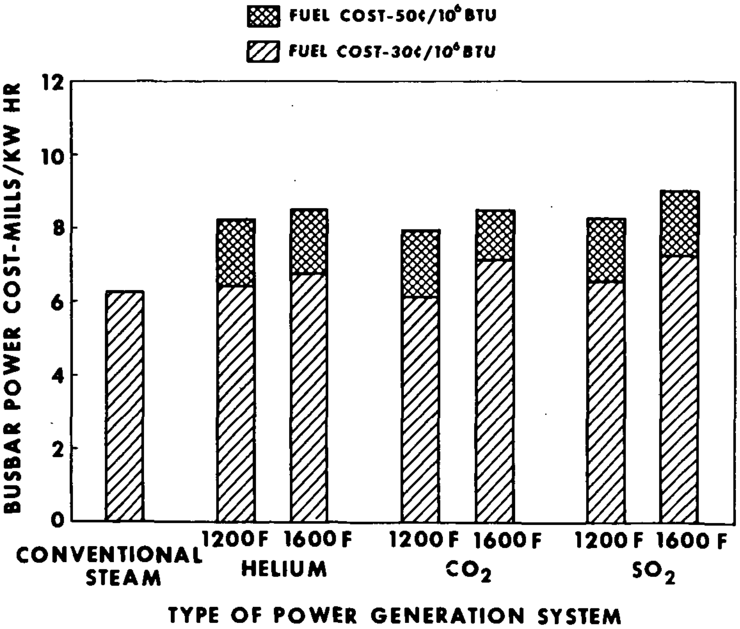


FIG. 7

PROJECTED DESIGN TECHNOLOGY FOR BASE-LOAD GAS TURBINE SYSTEMS

FUELS: METHANE (HHV = 1000 BTU/FT^3)
PRODUCER GAS (HHV = 162 BTU/FT^3)

PARAMETER	TIMF PERIOD		
	FIRST GENERATION (1970's)	SECOND GENERATION (1980's)	THIRD GENERATION (1990's)
TURBINE INLET GAS TEMPERATURE, F	1600 TO 2400	2000 TO 2800	2400 TO 3100
COMPRESSOR PRESSURE RATIO	8 TO 28	8 TO 36	8 TO 36
COMPRESSOR POLYTROPIC EFFICIENCY, %	89	92	93
TURBINE NOMINAL ADIABATIC EFFICIENCY, %	90	92	93
REGENERATOR AIRSIDE EFFECTIVENESS, %	60 TO 90	60 TO 90	60 TO 90
REGENERATOR TOTAL PRESSURE DROP, %	4 TO 8	4 TO 8	4 TO 8
TURBINE COOLING TECHNIQUE	IMPINGEMENT-CONVECTION	IMPINGEMENT-CONVECTION	1. IMPINGEMENT-CONVECTION 2. TRANSPIRATION
TURBINE COOLING AIR TEMPERATURE,F	1. SAME AS COMPRESSOR DISCHARGE AIR TEMPERATURE 2. PRECOOLED TO 125-TO-250 F		

FIG. 8

TURBINE INLET TEMPERATURE PROGRESSION FOR BASE-LOAD OPERATION

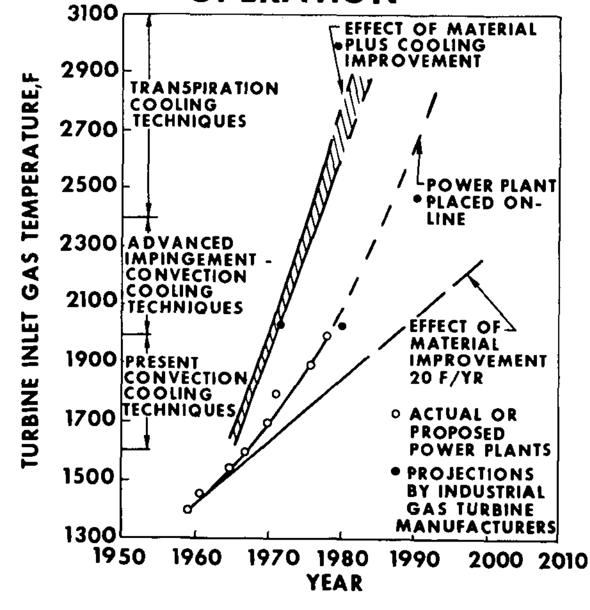


FIG. 9

EFFECT OF TURBINE COOLING FLOW ON BASE-LOAD SIMPLE-CYCLE GAS TURBINE PERFORMANCE

FUEL-METHANE (HHV = 1000 BTU/FT^3)
AMBIENT 80 F AND 1000 FT
FUEL DELIVERED TO BURNER AT COMPRESSOR OUTLET PRESSURE

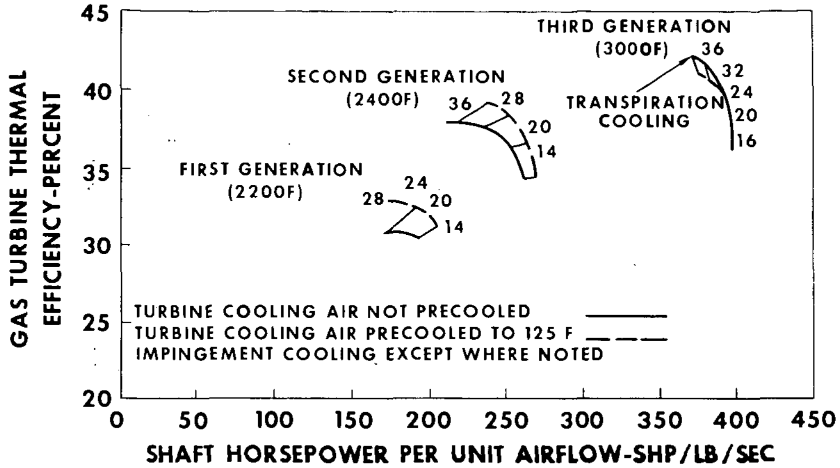


FIG. 10

THIRD GENERATION BASE-LOAD SIMPLE-CYCLE GAS TURBINE PERFORMANCE

FUEL-PRODUCER GAS (HHV = 162 BTU/FT^3)
AMBIENT-80 F AND 1000 FT
TURBINE COOLING CONFIGURATION :
ADVANCED IMPINGEMENT - CONVECTION

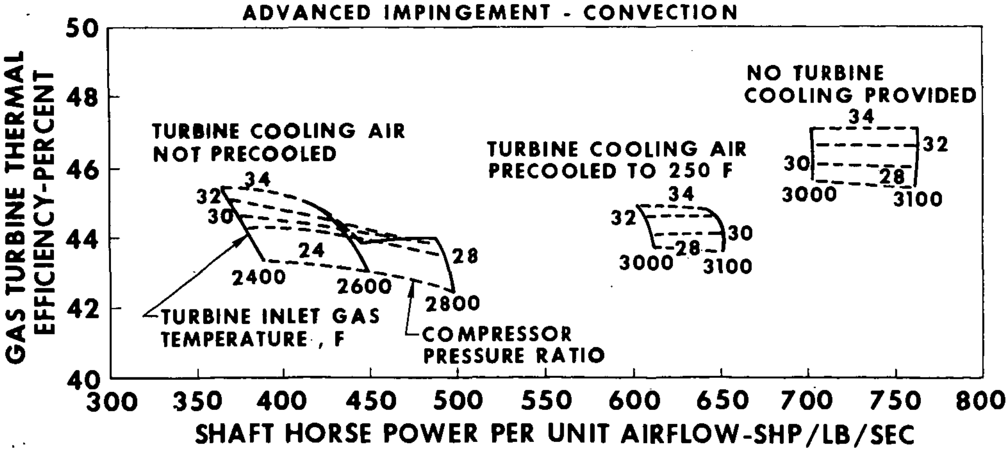


FIG. 11

COMBINED GAS-STEAM TURBINE SYSTEM

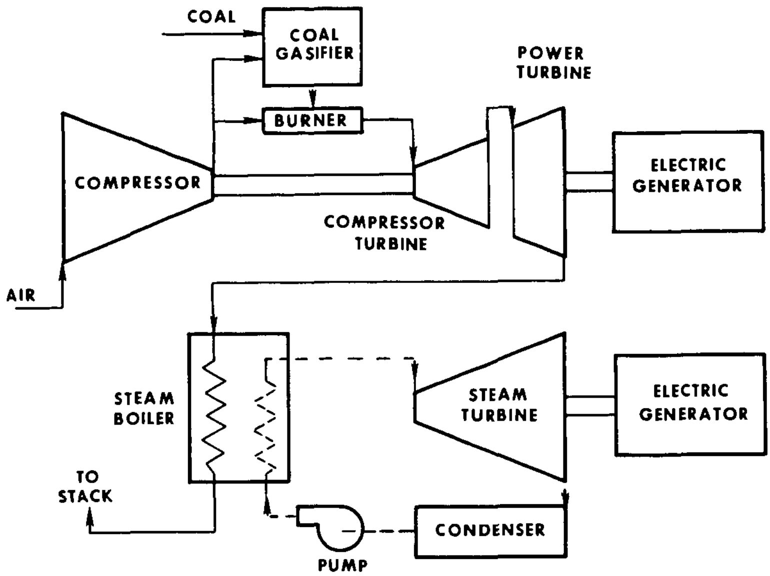


FIG. 12

PERFORMANCE OF EXHAUST-FIRED COMBINED SYSTEM

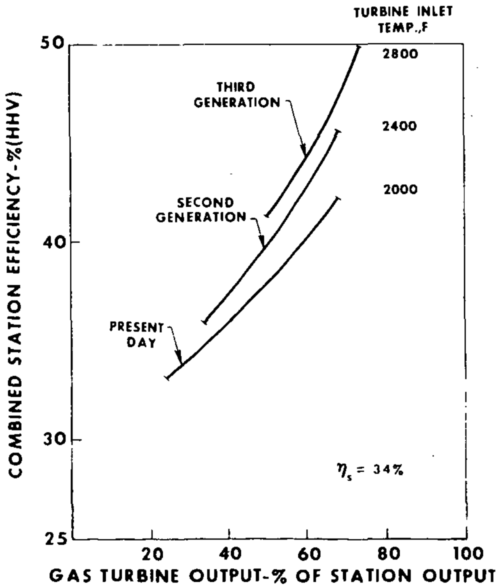


FIG. 13

PERFORMANCE OF UNFIRED WASTE HEAT COMBINED CYCLE

BASED ON USE OF 2400 PSIG/1000F/1000F STEAM CYCLE
EFFICIENCY = 38.8%
——— METHANE SUPPLIED AT BURNER PRESSURE AND 80F
– – – 162 BTU/FT³ GAS SUPPLIED AT BURNER PRESSURE AND 150F

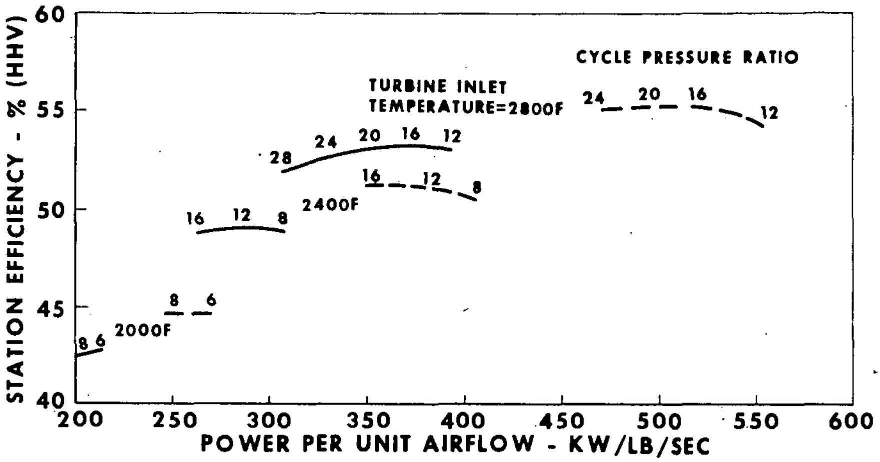


FIG. 14

PERFORMANCE OF UNFIRED WASTE-HEAT COMBINED CYCLE

FUEL - 162 BTU/FT^3 GAS SUPPLIED AT BURNER PRESSURE AND 150 F
STEAM CYCLE - 2400 PSIG/1000 F/1000 F, EFFICIENCY = 38.8%

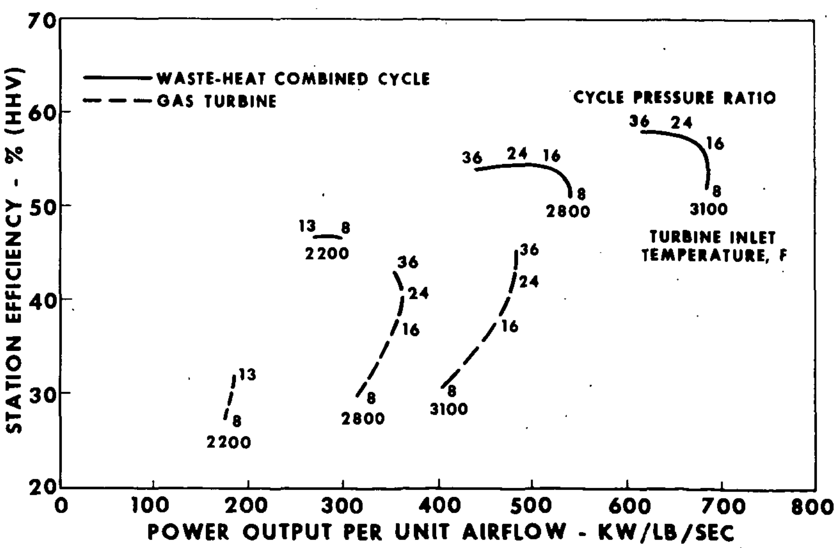


FIG. 15

EFFECT OF FUEL SUPPLY TEMPERATURE ON STATION PERFORMANCE

FUEL-162 BTU/FT^3 GAS SUPPLIED AT BURNER PRESSURE
STEAM CYCLE-2400 PSIG/1000 F/1000 F, EFFICIENCY = 38.8%
TURBINE INLET TEMPERATURE = 3100 F
CYCLE PRESSURE RATIO = 24

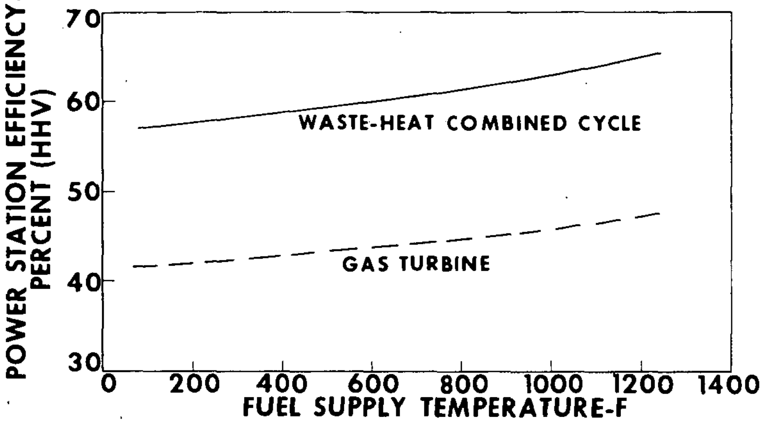


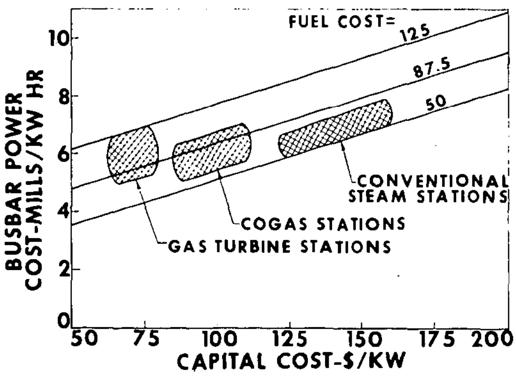
FIG. 16

BUSBAR POWER COSTS

1000-MW STATION
70% ANNUAL LOAD FACTOR

$$\text{POWER COST} = \frac{\text{CAPITAL CHARGES + FUEL COST + LABOR + MAINTENANCE}}{\text{KW-HR/YR}}$$

$$\text{FUEL COST} = \frac{\text{BASE FUEL COST}}{\text{STATION EFFICIENCY}}\ \text{¢}/10^6\ \text{BTU}$$



THE DETAIL DESIGN OF A 100-KILOWATT COAL-BURNING FUEL-CELL POWER PLANT

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ABSTRACT

A 100-kilowatt coal-burning fuel-cell process development plant has been conceived to provide technical and economic information for the design of a utility coal-burning fuel-cell power plant, to test fuel-cell battery performance and life under actual operating conditions, and to provide operating experience. The 100-kilowatt plant detail design incorporates an efficient combination of fuel-cell batteries and fluidized coal bed into a unique, flexible, modular, fuel-cell reactor unit. The reactor design achieves efficient transfer of heat generated during fuel-cell operation to the endothermic gasification of coal by inserting fuel-cell battery bundles into five inch diameter pipes located horizontally in a rectangular fluidized bed of coal. The fuel-cell battery bundles contain up to 40 fuel-cell batteries, 1/2 inch in diameter and approximately 30 inches long. Experimental and theoretical studies verify unique elements of the design.

The work recorded in this paper has been carried out under the sponsorship of the Office of Coal Research, U. S. Department of the Interior. N. P. Cochran and P. Towson have served as contract monitors. The Institute of Gas Technology, Dr. B. S. Lee and Harlan Feldkirchner, and Cameron Engineers, D. Lockwood, were subcontracted to prepare the detail design for the plant.

INTRODUCTION

Westinghouse, under contract to the Office of Coal Research, is developing solid-electrolyte fuel-cells which have the capability of producing electrical energy from coal at high efficiency in large scale power plants. The type of large scale, fuel-cell, power system envisioned comprises fuel-cell battery tubes, each containing many individual cells connected in series; a process for producing carbon monoxide and hydrogen fuel gases from coal; a means for cleaning and circulating the fuel gases; and a means for transferring the heat produced by the fuel cells to the endothermic coal gasification reactions between carbon and CO_2 and H_2O. The heat produced in fuel-cell operation is from resistive and polarization losses and the heat released in oxidation of the fuel gases. The design and operation of solid-electrolyte fuel-cells for application in power plants are reported elsewhere[1,2].

A coal-burning solid-electrolyte fuel-cell power system incorporating these components is shown in Figure 1. Coal is fed to a fluidized bed coal reactor along with a portion of the hot CO_2 and H_2O combustion products from the fuel-cell batteries. The coal reacts with those combustion products to form CO and H_2 fuel gases, which are recycled to the fuel-cell batteries after removal of particulate material and sulfur compounds. The fuel gas from the gas cleaning process is split into two streams - one going to a bank of fuel-cells where partial oxidation occurs and the other going to a bank where essentially complete combustion occurs. Gases from the second cell bank are discharged from the system while the partially oxidized fuel gases are recycled back to the coal reactor. The fuel-cell banks and the coal reactor are combined into a single unit to obtain maximum efficiency from the power plant. The potential advantages which this fuel-cell power system offer are:

- High efficiency (overall operating efficiencies near 60% are anticipated in full scale power plants),
- Minimum air and water pollution,
- Reduced plant size,
- Minimum cooling water requirements.

Westinghouse conceived the 100-kilowatt power plant as a means to test fuel-cell battery performance and life under actual operating conditions, to provide technical and economic information for the design of an economical and reliable coal-burning fuel-cell power plant, and to provide operating experience. The Institute of Gas Technology reviewed the concept and Cameron Engineers prepared the detail design of the 100-kilowatt plant.

BASIS

The design of the 100-kilowatt plant is based on the performance of a 100-watt generator[5] and recent data obtained using more economical electrode materials[1]. The fuel-cell banks are designed for fuel-cell battery tubes 1/2 inch in diameter and approximately 30 inches long. Approximately 40 fuel-cells would be connected in series on each battery tube. Projected fuel-cell operating characteristics for the 100-kW plant are:

1) average polarization voltage ≈ 0.15 volts

2) average fuel-cell resistance ≈ 0.15 ohms/cell

3) current density - varies from 300 ma/cm^2 in cell bank I to 50 ma/cm^2 at the end of cell bank II.

The fuel-cell operating characteristics were combined with the calculated thermodynamic open-circuit voltages to obtain the fuel-cell performance and efficiency. The average power produced per cell is approximately 0.22 watts /cell.

With these specifications, cell bank I requires approximately 6400 batteries and is designed to operate near 80% efficiency. Approximately 7000 batteries are required for cell bank II operating near 70% efficiency. Each bank is designed to produce approximately 60-kW of electrical power.

DETAIL DESIGN

Flow Diagram

A detail design for constructing a 100-kilowatt coal-burning fuel-cell process development plant has been completed except for the detail design of the fuel-cell battery bundle assemblies. The process flow sheet is presented in Figure 2. The material balance and the design are based on the Pittsburgh No. 8 char presented in Table I. Projected operating temperatures, pressures, and flow rates are given in Table II. Provision is made to record all char weights, stream temperatures, pressures, compositions, and flow rates necessary for material balances, energy balances, control, and general information on the performance of the plant.

The main reactor, which combines the fuel-cell batteries and fluidized coal bed, is the critical unit in the plant. The reactor is constructed of 10'1-1/2" x 5'2" x 15" rectangular modules which can be fabricated separately and bolted together. Overall dimensions of the reactor are shown in Figure 3. The fluidized coal bed is approximately 16 feet deep with a 6 foot disengaging section. Each of the 12 fuel-cell containing modules comprising the bed contains 40 pipes which are cantilevered into the reactor as shown in Figure 4. The horizontal pipes house the fuel-cell battery bundle assemblies. The fluidized coal bed for producing fuel gas for the cells surrounds the horizontal pipes. The arrangement facilitates efficient heat transfer from the fuel-cell batteries which operate around 1870°F to the fluidized coal bed which operates around 1750°F. The cantilevered pipes and long crossover pipes minimize stresses and provide for thermal expansion.

A representation of a fuel-cell bundle assembly in one of the horizontal pipes is illustrated in Figures 5 and 6. Fuel gas, rich in hydrogen and carbon monoxide produced in the fluidized coal bed surrounding the pipes, is fed to the inside of each fuel-cell battery. Air is supplied to the outside electrodes of the fuel-cell batteries. This is accomplished by feeding air to the end pipes in a module and allowing the air to pass to other pipes in the module through 1-1/2 inch crossover pipes which connect the 5 inch pipes. This is illustrated in Figure 4. The spent air is exhausted from the other end of the module and passed through a heat exchanger to preheat the inlet air as shown in the flow diagram.

TABLE I

CHAR FEED AND RESIDUE

Stream	Flow Rate, lb/hr	Composition, wt %	
Char Feed	69.16	C	70.50
		H	3.55
		O	10.54
		N	1.29
		S	3.46
		Ash	10.66
			100.00
Spent Char	12.04	C	34.88
		O	7.57
		N	0.57
		S	1.88
		Ash	55.10
			100.00

TABLE II
PROJECTED PLANT OPERATING CONDITIONS

Stream Number*	Description	T, °F	P, psig	Flow Rate lb/hr	Flow Rate lb moles hr.	Composition Mole %
1-1	Gasifier Effluent Recycle	1750	4.0	352.41	14.85	CO 68.20
1-2	Gasifier Effluent Recycle	1750	3.0	352.41	14.85	CO_2 7.50
1-3	Gasifier Effluent Recycle	923	3.0	352.41	14.85	H_2 19.85
1-4	Gasifier Effluent Recycle	800	3.0	352.41	14.85	H_2S 0.45
1-5	Gasifier Effluent Recycle	800	3.0	352.41	14.85	H_2O 3.40
1-6	Gasifier Effluent Recycle	800	3.0	352.41	14.85	N_2 0.60
						100.00
1-7	Gasifier Effluent Recycle	800	1.8	351.33	14.85	CO 68.20
1-8	Fuel Gas	800	1.8	351.33	14.85	CO_2 7.50
1-9	Fuel Gas	250	0.5	351.33	14.85	H_2 19.85
1-9A	Fuel Gas	250	25	--	--	H_2O 3.85
1-10	Fuel Gas	400	20	351.33	14.85	N_2 0.60
1-11	Fuel Gas	250	20	351.33	14.85	100.00
1-12	Fuel Gas	250	20	236.54	10.00	
1-13	Fuel Gas	250	20	114.79	4.85	
2-1	Fuel Cell Bank I Effluent Recycle	600	8.5	296.62	10.00	CO 41.30
2-2	Fuel Cell Bank I Effluent Recycle	1650	8.5	296.62	10.00	CO_2 34.40
						H_2 9.20
						H_2O 14.50
						N_2 0.60
						100.00

*Refer to Figure 2

TABLE II (Continued)
PROJECTED PLANT OPERATING CONDITIONS

3-1	Air	Ambient	0	1240	42.6	O_2	21.00
3-2	Air	70	20	1240	42.6	N_2	79.00
3-1A	Air	Ambient	20	--	--		100.00
3-3	Air	70	20	533.42	18.35		
3-4	Air	70	20	291.74	10.11		
3-5	Air	70	20	323.57	11.22		
3-6	Air	1455	13	291.74	10.11		
3-7	Air	1455	13	323.57	11.22		
4-1	Spent Air From Bank II	1870	7.5	256.96	9.13	O_2	3.00
4-2	Spent Air From Bank II	200	2.0	256.96	9.13	N_2	97.00
							100.00
5-1	Spent Air From Bank I	1870	7.5	231.68	8.24	O_2	3.00
5-2	Spent Air From Bank I	200	2.0	231.68	8.24	N_2	97.00
							100.00
6-1	Bank II Spent Fuel	600	8.5	181.38	4.85	CO	1.70
6-2	Bank II Spent Fuel	250	2.0	181.38	4.85	CO_2	74.00
						H_2	0.50
						H_2O	23.20
						N_2	0.60
							100.00
7-1	Startup Gas						

Two 21.5 ft^3 char feed hoppers are used to feed the coal to the system. The first hopper is a storage unit for introducing char into the system. The second hopper is the feed hopper.

Char, sized to minus 30 mesh, is screw fed to the center module. This module contains horizontal pipes in order to maintain uniform fluidization throughout the bed and the pipes are also designed to serve as auxiliary gas distributors for start-up. The fuel gas produced in the fluidized bed reactor passes through an electrically traced line to a cyclone. Dust is collected for analysis and is not returned to the bed. The gas then passes through a recycle heat exchanger where heat is supplied to the spent fuel from cell bank I before it returns to the reactor. The fuel gas is cooled to approximately 800°F in a guard cooler before final particulate removal in an electrostatic precipitator. Sulfur is removed to a controlled level by a zinc oxide absorber. Two absorbers are provided to permit zinc oxide replacement during extended runs. The units are 2 ft diameter and 6 ft high and are designed to operate for approximately 3 days with a sulfur concentration of 8-10 grains/100 SCF. The hydrogen sulfide concentration in the gas leaving the absorbers is monitored and the hydrogen sulfide concentration controlled. This capability is provided since trace amounts of hydrogen sulfide may be required to inhibit carbon deposition as the gas is reheated through the critical deposition range. A 2 ft diameter by 6 ft high water gas shift reactor is provided to increase the hydrogen content of the gas. Conversion of H_2O in the fuel gas to H_2 is desirable since H_2 minimizes the polarization voltage losses in the fuel-cells. The fuel gas is compressed through a 20 hp compressor before being split into separate streams and fed to the two fuel-cell banks.

The fuel gas is split into two unequal portions. Approximately one-third of the gas goes to cell bank II and two-thirds to cell bank I. The fuel gas to cell bank II is approximately 97% oxidized and the resulting CO_2-H_2O gas is exhausted from the system. The fuel gas to cell bank I is only partially oxidized (see Table II) and is recycled to the fluidized coal bed. Upon leaving cell bank I, the gas is preheated through the recycle heat exchanger and then introduced into the reactor through the gas distribution module. The gas distribution module, shown in Figure 7, consists of a single row of ten 5 inch diameter pipes which extend across the reactor with 0.0576 inch holes in two parallel rows along the bottom of each pipe. Spent char is removed from the reactor by a residue discharge screw below the gas distributor.

In order to achieve early operation and to facilitate gaining operating experience, the system is designed to operate without fuel-cell batteries by installing electric heaters in the horizontal pipes in place of fuel-cells. When operating with electric heaters the gas normally sent to cell bank II is vented. The composition of the remaining gas can be adjusted by simulating the oxidation process with an oxidizer or by using make-up gas. Thus, the gas can be returned to the fluidized bed coal reactor simulating operation with fuel-cell batteries. This procedure enables the system to be checked out before installing the valuable fuel-cell batteries.

Cost Estimate

The cost estimate for the construction of the 100-kilowatt process development plant, excluding fuel-cell batteries and the bundle assemblies, is $2.0 million. A summary of the cost breakdown is presented in Table III.

TABLE III

COST ESTIMATE SUMMARY

Description	Equipment, Materials, and Labor	Total
Purchased Equipment	$131,000	
Fabricated Equipment	179,000	
Instruments and Control Panel	297,100	
Electrical	152,800	
Structural and Concrete	207,800	
Valves	93,900	
Pipe and Tubing	156,300	
Insulation and Refractory	26,300	
Sub-Total		$1,244,200
Indirect Costs (37.6% of equipment, materials and labor)		467,800
Sub-Total		1,712,000
Contingency (15%)		257,800
Total Estimated Plant Cost		$1,969,800

DISCUSSION

Design Features

The 100-kilowatt process development plant is designed to test and prove out fuel-cell batteries and to provide technical and economic information for scale-up. The process development plant design cannot be directly scaled to an economic commercial design. Much longer fuel-cell batteries are required for a power generator serving as a central station plant -- the 30 inch battery length currently limits the reactor size and shape severely. Batteries on support tubes which can be directly immersed in the fluidized coal bed are also required -- the costs of separate fuel-cell protection tubes and manifolding appear prohibitive for a central station plant. However, answers required for the development of an economical and reliable coal-burning fuel-cell power system can be provided by the process development plant. The plant can be used to study:

1) fuel-cell battery performance and life at actual operating conditions;

2) procedures for operating large numbers of fuel-cell batteries; e.g., methods to vary power loads, methods of current collection, requirements for abnormal operating conditions such as cell failure or short circuits, etc.;

3) tolerance levels for dust and hydrogen sulfide for the fuel-cell batteries;

4) reaction rates of different chars and coals at various operating conditions;

5) heat transfer between the fuel-cell batteries and the fluid bed;

6) conditions of fluidization;

7) materials of construction;

8) start-up, shut-down, and emergency procedures;

9) maintenance and replacement problems; and

10) safety requirements.

In order to carry out these studies, three phases of plant operation are proposed:

1) operation of the coal reactor system using electric heaters in place of fuel-cells.

2) study of test fuel-cell bundles to evaluate mechanical and electrical operability under operating conditions and the testing of the fuel and air piping systems,

3) operation of the system as a whole to produce 100-kW.

The first two steps are recommended in order to check out and characterize the system before inserting the valuable fuel-cell batteries. A summary of the projected experimental program is presented in the Appendix.

The 100-kilowatt process development plant design incorporates an efficient combination of fuel-cell batteries and fluidized coal bed into a unique, flexible fuel-cell reactor unit. The reactor is designed to accept 20 inches of active battery length; however, the modular design permits modules to be constructed which would accept horizontal fuel-cell batteries up to 3-1/2 feet long. The reactor could also be modified to accept batteries in the vertical position greater than 10 feet long. The reactor volume is based on conservative estimates of the gasification rate[3,4,6]. This will permit a wide range of coals, cokes, or chars to be studied. The reactor can also be used to study simultaneous fluidized bed coal gasification and desulfurization with limestone sorbents, which may be attractive for commercial application. When fuel-cell performance reaches 0.5 watts/cell, it will be possible to produce more than 300 kilowatts in the present system. The flexibility of this reactor design far outweighs any considerations for going to a compact design at this stage of the development. The system does not give high overall system efficiency, since it is not practical to eliminate all the heat losses on the 100-kilowatt development unit. However, the design does provide the information for projecting the efficiency of large-scale plants.

The auxiliary equipment has been designed using present day technology. This will minimize start-up and operating difficulties and will allow the evaluation of problems associated with the development of a commercial power system.

Evaluation of Critical Design Features

The reactor, which combines the fuel-cell batteries and fluidized coal bed, is the critical unit in the plant. In order to verify the operability of the design, Westinghouse conducted experimental and theoretical studies and had IGT and Cameron Engineers make a thorough evaluation of the structural design. Results of these evaluations indicate the reactor design is structurally sound and operable.

Fluidization and heat transfer experiments were made on a Plexiglas model of the reactor[7,8]. Temperature profiles were also recorded around a 1 inch diameter tube with internal heat generation in a 3-9/16 inch diameter fluidized bed of char maintained at 1600°F[9]. The results of these studies were combined with the projected fuel-cell performance to evaluate the fuel-cell reactor system[10]. The analysis indicates the maximum temperature gradient between the fluidized bed and the fuel-cell bundles will be less than 150°F and the maximum temperature gradient within the bundles will be less than 100°F. The models predict the temperature gradient will be approximately 100°F and 50°F, respectively with uniform heat transfer from the fuel-cell bundle assemblies to the fluidized bed. These results indicate the fluidized bed will operate at 1700°F or higher where the reaction kinetics are favorable without exceeding the upper temperature limit of the fuel-cell batteries.

The horizontal pipes in the reactor are subjected to a reducing gas containing hydrogen sulfide near 1800°F. In order to assure reactor life and the ability to evaluate different coals and chars, several materials were considered and tested for the horizontal pipes. The results of corrosion tests indicate that Incoloy 800 can be used with de-sulfurized chars (< 0.7% sulfur in coal or char). Corrosion tests with L-605, a cobalt base alloy, indicate sulfur contents in coals or chars near 3% can be used based on a 10,000 hour reactor life. L-605 has greater strength and a smaller thermal expansion than Incoloy 800 and can be substituted in the design without any modifications.

Many aspects of the reactor design cannot be completely evaluated until the unit is built; such as gas distribution, reliability of welds and the air piping design, fluidizing conditions, etc. All of these aspects have been evaluated on bench scale apparatus or analyzed to check the design.

CONCLUSIONS

A 100-kilowatt coal-burning fuel-cell process development plant has been designed. The plant will test fuel-cell battery performance and life under actual operating conditions; will provide technical information on gasification, heat transfer, coal handling, materials, and control; and will provide operating experience. The design provides an efficient combination of the fuel-cell batteries and the fluidized coal bed into a unique, flexible, modular fuel-cell reactor unit. This is achieved by inserting fuel-cell battery bundle assemblies horizontally into a rectangular fluidized bed of coal. Experimental and theoretical evaluations of the fuel-cell reactor unit indicate the design is structurally sound and operable. The estimated cost for constructing the 100-kilowatt process development plant is $2.0 million.

The detail design of the process development is complete. The decision on whether to build the plant has been deferred while additional development is being carried out on the batteries.

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APPENDIX

PROJECTED EXPERIMENTAL PROGRAM FOR 100-KILOWATT PROCESS DEVELOPMENT PLANT

I. OPERATION WITHOUT FUEL-CELL BATTERIES

A. Fluidization

1. Cold Studies

   Fluidize the bed with an inert to evaluate minimum fluidizing velocity, ΔP, uniformity of fluidization, cyclone operation, dust sampling systems, effect of particle size distribution, operation of gas distributor, coal feed system, ash removal system, compressors, level control, and auxiliary equipment.

2. Hot Studies

   Fluidize coal bed with inert gas and supply heat with electrical heaters. Evaluate operation of equipment as suggested during cold study.

B. Start-Up and Shut-Down Procedures

1. Check start-up procedure without fuel-cells present - combustion of char, heating rates, operation techniques.
2. Check emergency and shut-down procedures.

C. Heat Transfer

1. Study with Inert Gas

   Study effect of temperature, flow rates, particle size on heat transfer from the fuel-cell pipes. Determine the effects of fuel-cell pipes not producing heat.

2. Study with Fuel Gas

   In addition to the parameters investigated with the inert gas, determine the effect of various coals.

D. Reaction Rate

1. Study the effect of temperature, gas flow rate, oxygen content of the inlet gas, H/C ratio in the gas and solid feed streams, char residence time, particle size, and char composition on rates of gasification.
2. Study segregation of ash in the bed, and the effects of various ash contents and compositions on bed performance.

E. Sulfur Removal

1. Study removal of sulfur in the fluidized bed.

II. OPERATION WITH TEST FUEL-CELL BATTERY BUNDLES

Note: The following evaluations will be conducted on dummy fuel-cell bundles (i.e., a bundle of ceramic support tubes) before operating with production fuel-cell bundles.

1. Check manifolding, instrumentation, and control of the gas flow to the batteries.
2. Determine the effect of vibrations on the batteries, inlet gas temperatures and flow rates on cell performance, pressure drop across the batteries, various fuel gas compositions from the reactor on cell performance.
3. "Short" life tests of battery bundles.

III. 100-KILOWATT OPERATION

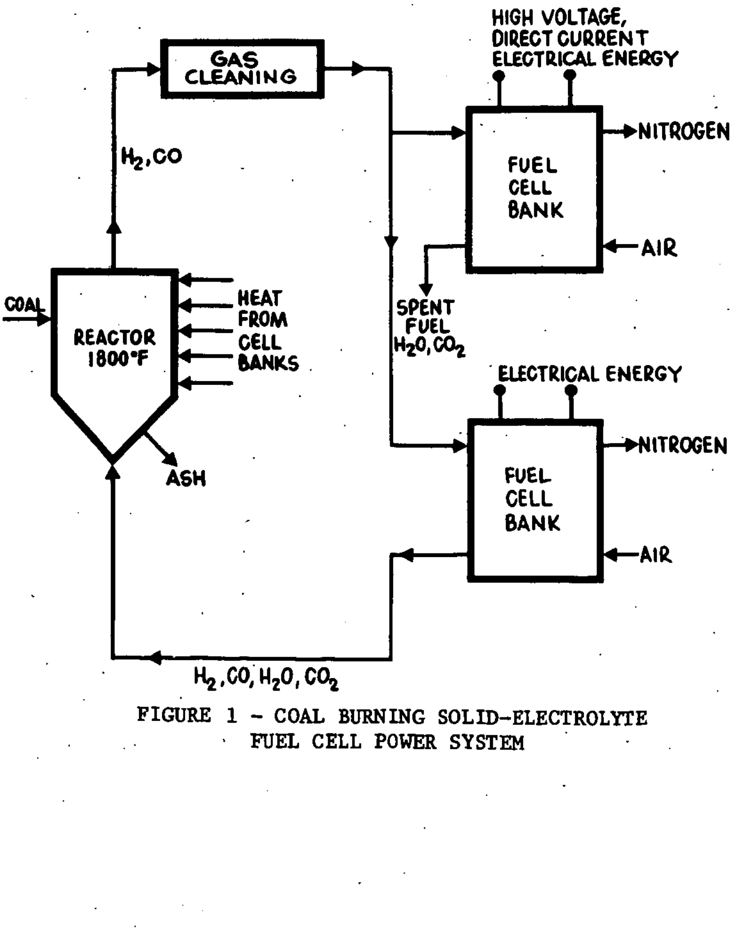

FIGURE 1 - COAL BURNING SOLID-ELECTROLYTE FUEL CELL POWER SYSTEM

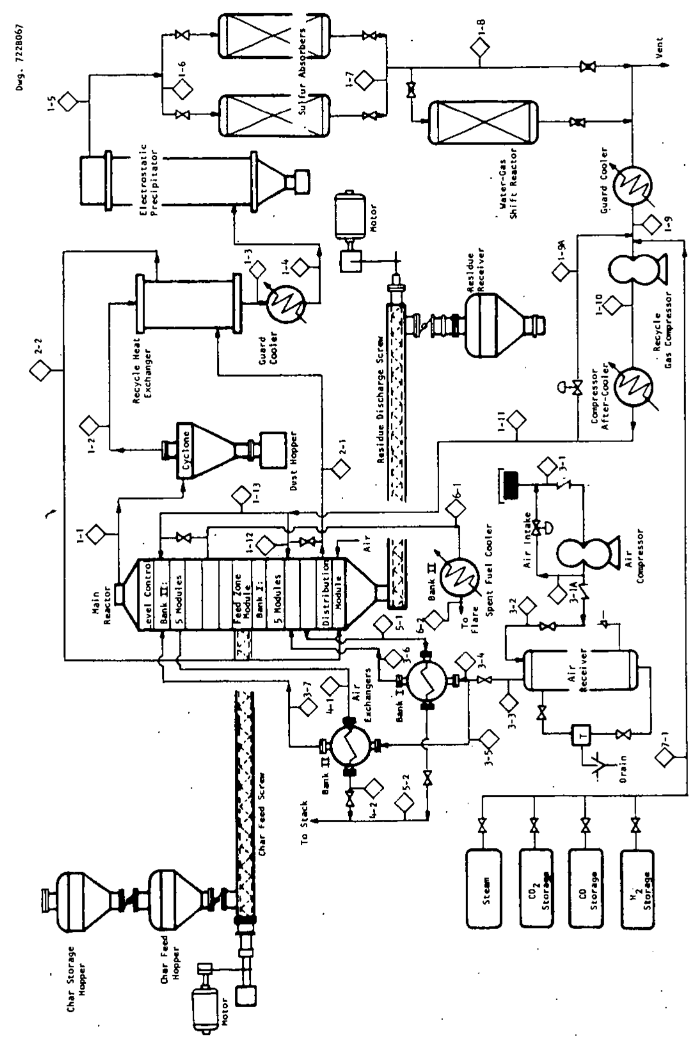

FIGURE 2 - PROCESS FLOW SHEET FOR 100-KW PROCESS PLANT

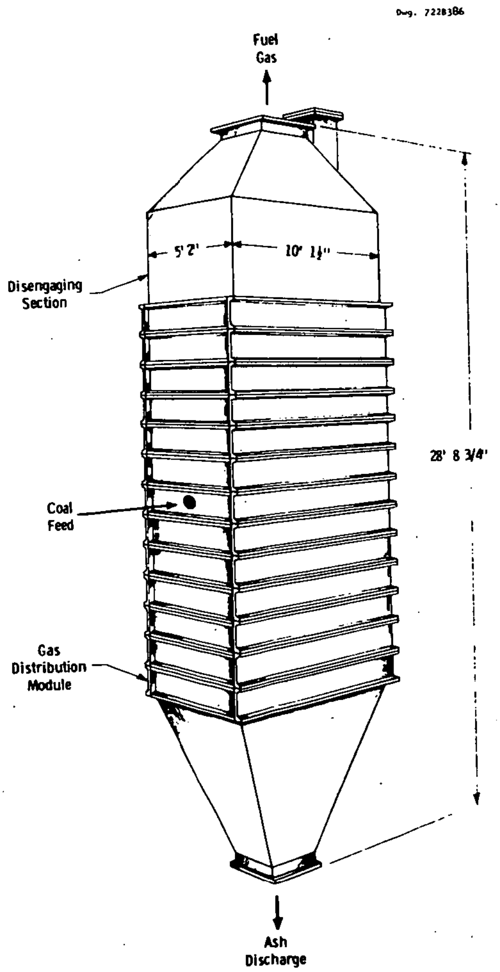

FIGURE 3 - 100-KILOWATT COAL-BURNING FUEL-CELL REACTOR

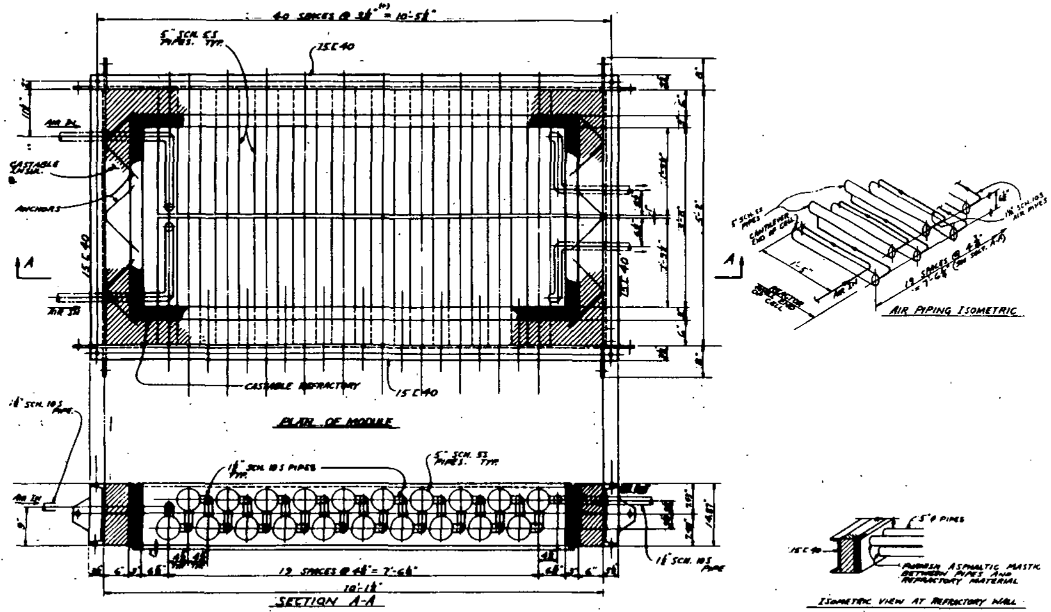

FIGURE 4 - FUEL-CELL REACTOR MODULE

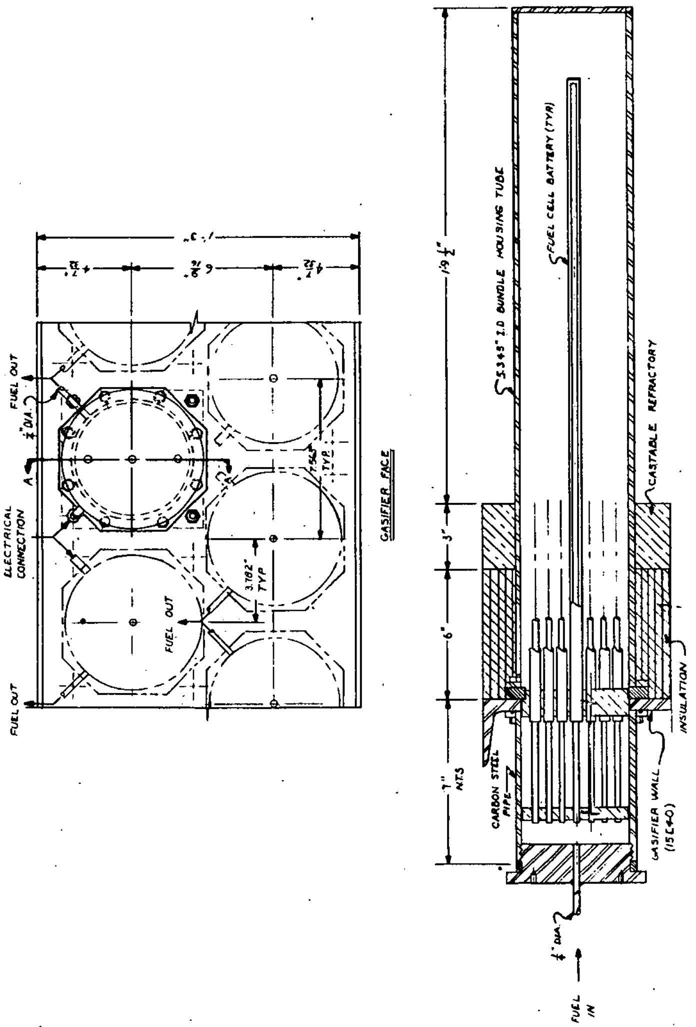

FIGURE 5 - FUEL CELL BUNDLE ASSEMBLY

Dwg. 855A849

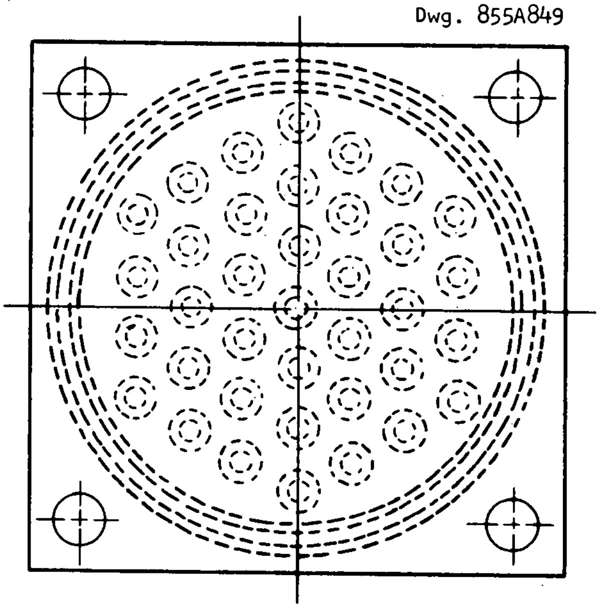

FIGURE 6 - END VIEW OF FUEL CELL TUBE BUNDLE ASSEMBLY

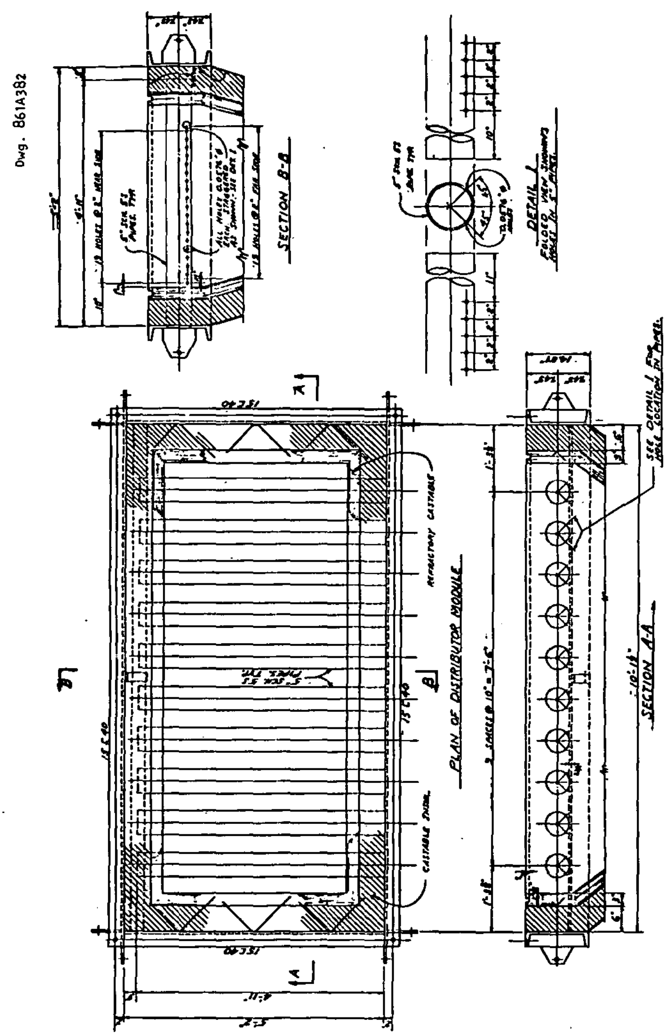

FIGURE 7 - GAS DISTRIBUTION MODULE SCHEMATIC